SNOX DEMONSTRATION PROJECT

FINAL REPORT VOLUME II: PROJECT PERFORMANCE AND ECONOMICS

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DISCLAIMER

ABSTRACT

The SNOX process, developed by Haldor Topsoe A/S and demonstrated and marketed in North America by ABB Environmental Systems (ABBES), is an innovative process which removes both sulfur dioxide and nitrogen oxides from power plant flue gases. Sulfur dioxide is recovered as high purity, concentrated sulfuric acid and nitrogen oxides are converted to nitrogen gas and water vapor; no additional waste streams are produced. As part of the Clean Coal Technology Program, this project was demonstrated under joint sponsorship from the U.S. Department of Energy, Ohio Coal Development Office, ABBES, Snamprogetti, and Ohio Edison.

The project objective was to demonstrate the SO_2/NO_x reduction efficiencies of the SNOX process on an electric power plant firing high-sulfur Ohio Coal. A 35-MWe demonstration has been conducted on a 108-MWe unit, Ohio Edison's Niles Plant Unit 2, in Trumbull County, Ohio. The $31.4 million project began site preparation in November 1990 and commenced treating flue gas in March of 1992. A parametric test program has been completed.

The following Project Performance and Economics Report has been prepared for Phase III of the SNOX Demonstration Project as described in Co-Operative Agreement No. DE-FC22-90PC89655 dated December 20, 1989. The report presents a description of the technology, results from the 33 month testing and operation phase, and information from a commercial scale economic evaluation.

During the demonstration, the process met or exceeded its design goals of 95% SO_2 removal, 90% NO_x removal, and production of commercial grade (>93.2 wt.%) sulfuric acid. The plant was operated for approximately 8000 hours and produced more than 5600 tons of acid, which was purchased and distributed by a local supplier to end users.

Projected economics for a 500 MWe commercial SNOX plant indicate a total capital requirement of 305 $/kW, levelized incremental cost of power at 6.1 mills/kWh, 219 $/ton of SO_2 removed, and 198 $/ton of SO_2+NO_x removed (all at constant dollars).

POINT OF CONTACT

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EXECUTIVE SUMMARY

The following Project Performance and Economics Report has been prepared as part of Phase III of the SNOX Demonstration Project as described in Co-Operative Agreement No. DE-FC22-90PC89655 dated December 20, 1989. The report presents a description of the technology, results from the 33 month testing and operation phase, and information from a commercial scale economic evaluation.

The SNOX Demonstration Project utilizes a highly efficient catalytic process that removes SO_2 and NO_x from flue gases and generates salable sulfuric acid. The integrated design of the process enables high removal efficiencies, no waste production, and increased thermal efficiency of the boiler. The Demonstration Plant is located at the Ohio Edison Niles Power Plant near Niles, Ohio, situated on 130 acres along the southern bank of the Mahoning River. This power station is part of the Ohio Edison System, which serves approximately 9,000 square miles in central and northeastern Ohio and western Pennsylvania. One-third of the flue gas from the Niles Station Unit 2 boiler is treated in the process.

One of the sixteen projects selected for funding under Round II of the Clean Coal Technology Program was the SNOX process demonstration proposed by Asea Brown Boveri Environmental Systems (ABBES). The total project cost was projected to be $31.4 million with the co-funders being: DOE ($15.7 million); OCDO ($7.8 million); ABBES and Snamprogetti ($6.7 million); and Ohio Edison ($1.2 million). The project was selected on September 28, 1988 and the Cooperative Agreement was signed on December 20, 1989.

The execution of the SNOX Demonstration Project was divided into three phases which span approximately sixty months. These phases are identified as follows:

- Phase I: Design and Permitting
- Phase IIA: Long Lead Procurement
- Phase IIB: Construction and Start-Up
- Phase III: Operation, Data Collection, and Reporting

Site preparation and installation of foundations began in November 1990 and construction was completed in November of 1991. Equipment commissioning was conducted following completion of construction and the system was first operated on flue gas in March of 1992. The project completed operation and testing under Phase III in December of 1994.

Originally the overall program was 48 months in length and was scheduled to end in December of 1993. A task for Site Restoration was included in Phase II-B funding in the event that Ohio Edison did not opt to retain the plant. During the second half of 1993, Ohio Edison announced that it would retain the plant and funds that were designated for dismantling were reapportioned into the operating phase of the program for testing and system modifications.

In order to demonstrate and evaluate the performance of the SNOX process during the Clean Coal Technology Program, operating data were collected and parametric tests conducted to characterize the process and equipment. The primary objectives for the SNOX Demonstration Project were as follows:

- Demonstrate NO_x and SO_2 removals of 90 and 95%, respectively
- Demonstrate the commercial quality of the product sulfuric acid.
- Satisfy all Environmental Monitoring Plan requirements.
- Perform a technical and economic characterization of the technology.

The demonstration project served to demonstrate the high performance of the SNOX technology in the North American power generation industry, i.e. with U.S. fuels and operating staff. The first and foremost objective of the project was to successfully apply the technology and proprietary equipment to a power plant firing high-sulfur coal so as to confirm the capability to economically meet the pollution control needs of that market. Several supporting objectives aided in meeting this primary objective. These supporting objectives were to confirm the results achieved at a Danish pilot facility, demonstrate the marketability and economic credits of the sulfuric acid and heat energy by-products, confirm the estimated low operating and maintenance costs, and define any limitations of each piece of major equipment with respect to a utility environment.

The SNOX technology consists of four key process areas, which are NO_x reduction, SO_2 oxidation, sulfuric acid (H_2SO_4) condensation and acid conditioning. NO_x removal is achieved by selective catalytic reduction (SCR), and SO_2 removal is achieved by catalytic oxidation to SO_3. The SO_3 in turn is reacted with water to form H_2SO_4. The key to the recovery of H_2SO_4 is a proprietary glass-lined, falling film condenser.

Total cost for the initially designed demonstration plant was $20.0 million. Additions and modifications made to the plant during the program brought the final cost to $20.9 million.

A test program was conducted in support of the stated objectives as follows:

1. Execute parametric test batteries on all major pieces of equipment.
   - Fabric filter
   - SCR system
   - SO_2 converter
   - WSA condenser
   - Gas/gas heat exchanger
   - Catalyst screening unit
2. Quantify process consumptions.
   - Power
   - Natural gas
   - Catalysts
   - Cooling water
   - Potable water
   - Ammonia
3. Quantify process productions.
   - Sulfuric acid
   - Heat
4. Quantify personnel requirements.
5. Evaluate all materials of construction.

Sulfur dioxide removal was consistently in the 95 to 96% efficiency range, and nitrogen oxides removal exceeded its target value by 3 - 4 percentage points, typically being 93 to 94% efficiency. With respect to sulfuric acid quality, its concentration and composition have met or exceeded the requirements of the Federal Specification for Class 1 for species analyzed. Commercial grade acid is specified as 93.2 wt.% and the demonstration plant acid was consistently in excess of this value.

The plant was operated for approximately 8000 hours and produced more than 5600 tons of acid.

A local acid supplier was contracted to purchase and distribute the acid from the plant once operation began. The company is a large regional marketer and producer of sulfuric acid serving the industrial Midwest in New York, Ohio, Michigan and Illinois. The acid is being sold primarily to the agricultural industry and to the steel industry for pickling.

An economic case study was performed for the SNOX process for a commercial size unit. A 500 MWe unit was assumed for the study, firing 3.2% sulfur coal. Total capital requirement for the plant is 305 $/kW. The levelized incremental cost is 6.1 mills/kwh on a constant dollar basis or 7.8 mills/kwh on a current dollar basis. The equivalent costs per ton of SO_2 removed are $219/ton on a constant dollar basis and $284/ton on a current dollar basis.

Comparison of the SNOX process economics with those of competing technologies shows advantages for SNOX, particularly when a market exists for the by-product H_2SO_4.

1.0 INTRODUCTION

1.1 Purpose of the Project Performance and Economics Report

The primary purpose of the Project Performance and Economics Report (PPER) is to provide a technical account of the total work performed under the SNOX Demonstration Project cooperative agreement. It is a comprehensive description of the results achieved, technical readiness, and ABB Environmental System's (ABBES) view and plan for commercialization of the technology demonstrated. The PPER, along with Volume I, the Public Design Report, serves as the main reference for parties interested in the technology: to determine the achievements of the project and to assist them in assessing the technical and economic applicability of the technology to their particular situations.

The PPER discusses the background of the project, any changes to the design made after the Public Design Report was issued, the technical accomplishments, the process economics, the environmental performance, and the applicability of the demonstrated technology to the available market. This information is provided to assist the private sector in judging commercial potential and making informed decisions on commercial readiness. Additionally, this information will assist federal, state, and local authorities in making sound policy and regulatory decisions regarding commercial deployment of the clean coal technology covered in the report.

1.2 Overview of the Project

1.2.1 Background and History of the Project

The SNOXJ process combines two Haldor Topsoe technologies, i.e. the innovative WSA process for the removal and recovery of sulfur dioxide as concentrated sulfuric acid (H_2SO_4) and the Selective Catalytic Reduction (SCR) of nitrogen oxides. The first plant using the principles of the WSA, or Wet-gas Sulfuric Acid, process without SCR was commissioned in 1963 in Lacq, France. This plant, which used a conventional acid absorption tower with circulating sulfuric acid

(WSA-1 process), treated a dust-free off-gas containing 1% SO_2. In 1980, two additional WSA-1 plants, treating off-gas with 10-15% hydrogen sulfide (H_2S), were commissioned in Sweden.

Limitations of the conventional acid tower led to the design of the WSA condenser and the development of the WSA-2 process which has replaced the WSA-1 process. The condenser is an air-cooled falling-film type of unique design. The flue gas flows through vertical glass tubes and by careful control of the temperature difference across the gas film inside the tubes and of the turbulence of the gas, sulfuric acid is condensed essentially without formation of acid mist.

The first industrial WSA-2 plant, cleaning 7,800 scfm (12,000 Nm^3/h) of off-gas from a molybdenum roaster in Sweden, has been in operation since May of 1986. The plant treats off-gas with 0.5-1.5% SO_2 at a removal efficiency of >95% and is recovering 95-96% sulfuric acid. Another WSA-2 plant designed to treat 78,000 scfm (125,000 Nm^3/h) of off-gas from a pulp mill in Taiwan was started up in 1990. Presently, 17 WSA-2 plants serving a variety of industries are in operation or under construction.

Between 1983 and 1985, Haldor Topsoe developed a catalyst for the denitrification of flue and exhaust gases. The first pilot testing of this catalyst was performed on diesel exhaust from a stationary engine on the Faroe Islands. The WSA-2 process was then combined with this de-NO_x technology to form the SNOX process for simultaneous removal of NO_x and SO_2. The SNOX process was first tested in December of 1985 at the Amagervaerket Power Station, Copenhagen, Denmark, with a 62 scfm (100 Nm^3/h) bench-scale plant.

In November, 1987, a 3 MWe (6,200 scfm, 10,000 Nm^3/h) demonstration SNOX plant was started up at Skaerbaekvaerket, Skaerbaek, Denmark. Intended to provide process data for the engineering of full-scale coal-fired utility power boilers, this pilot plant was designed so that all critical equipment, such as the bag filter, catalytic reactors, and WSA condenser, had the same modular sizes as in a full-scale plant. The WSA condenser, for instance, used the same glass tubes, tube pitch, construction materials, and details of construction as would a larger unit. In 1991, having fulfilled its purpose, this facility was decommissioned.

In 1989 a contract for the retrofitting of an existing power plant (305 MW) with the SNOX process was signed with a Danish power producer, ELSAM. This plant was officially commissioned in November of 1991 and has been routinely operating at design capacity (620,000 scfm, 1,000,000 Nm^3/h). All guarantees have been satisfied, with 95% removals of SO_2 and NO_x achieved. This station received the 1992 International Powerplant Award from Electric Power International Magazine for its balance of both energy and environmental needs.

Also in 1991, a demonstration scale SNOX plant was commissioned by Snamprogetti S.p.A which treats 62,000 scfm (100,000 Nm^3/h) of flue gas from a petroleum coke (6% sulfur) fired power plant owned by Enichem S.p.A. in Italy. This plant has met or exceeded all process design objectives and is supplying sulfuric acid to an adjacent petrochemical complex for fertilizer production. Removals of SO_2 and NO_x are greater than 96% at this facility.

In late 1989, a contract with the U.S. Department of Energy (DOE) was signed as part of the Clean Coal Technology (CCT) Program, Round II, to demonstrate the SNOX process at a U.S. utility power station. The host site was the Niles Power Plant of the Ohio Edison System. This project served to demonstrate the high performance of the SNOX technology in the North American power generation industry, i.e. with U.S. fuels and operating staff. The project also proved the commercial quality of the sulfuric acid produced by the SNOX process in the U.S. marketplace. Including the U.S. plant, Topsoe has supplied 6 SNOX plants to date.

1.2.2 Project Organization

Total project cost for the SNOX Demonstration was estimated to be $31,438,408. The co-funders were the DOE ($15,719,200), the Ohio Coal Development Office (OCDO - $7,859,602), ABBES along with Snamprogetti ($6,665,606), and the Ohio Edison Company ($1,194,000). The project was selected by the Source Selection Official on September 28, 1988, and assigned to the Pittsburgh Energy Technology Center (PETC). A multi-discipline team was assembled at PETC and ABBES respectively, which consisted of the project manager, a contract specialist, legal counsel, an environmental coordinator, a cost/price analyst, a post-award auditor, a patent counsel, a business financial advisor, and a public relations person. The project manager coordinated the activities of the team members throughout the life of the project.

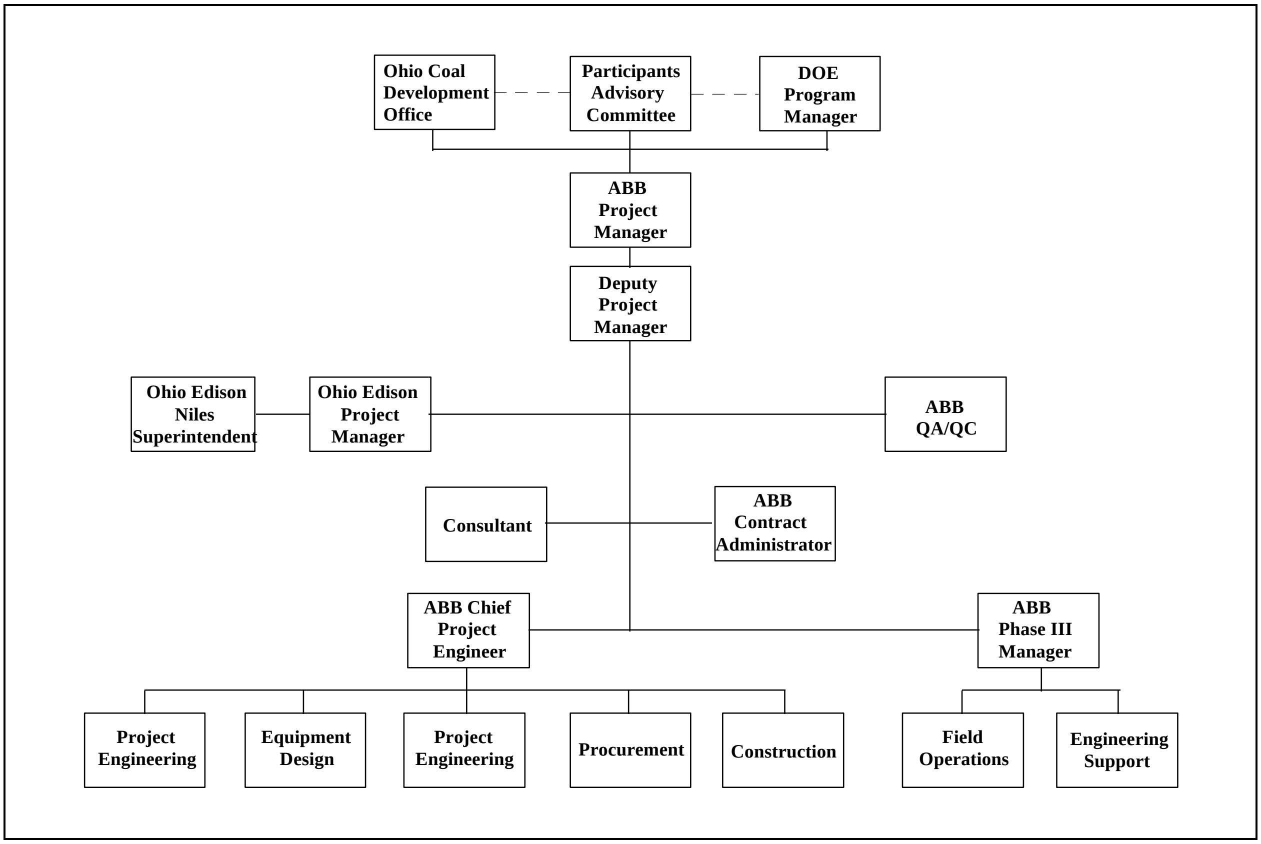
The project was managed by the Participant's (ABBES) Project Manager, who was the principal contact with DOE regarding the administration of the agreement. Assisting the ABBES Project Manager was a Deputy Project Manager from Snamprogetti who ensured that the design, procurement and construction work fulfilled the requirements of the SNOX technology.

A Participants Advisory Committee was formed and is comprised of personnel from ABBES, Snamprogetti, DOE, Ohio Edison, and OCDO. This Committee meets as needed to review the project, assess plans and provide advice on correcting any deficiencies. The Participants Advisory Committee is intended to be working group of personnel directly involved in the project and ensures that the objectives of each participating organization are met. The Participants Advisory Committee does not direct ABBES.

ABBES was responsible for all aspects of project performance under this Cooperative Agreement as set forth in the Statement of Work. The ABBES Project Manager was the authorized representative for the technical and administrative performance of all work performed under this Cooperative Agreement. He was the single authorized point of contact for all matters between ABBES, DOE, andco-sponsors.

The Deputy Project Manager from Snamprogetti worked closely with the ABBES Project Manager to assure that program activities are conducted within the schedule and operating requirements of the plant.

The Project Team also included a Project Engineer, who reported to the Project Manager and was responsible for all technical work, designation of work packages, engineering schedules, drawing submittals, field liaison and direction of all design work. Also, he was responsible for overseeing the Lead Discipline Engineers.

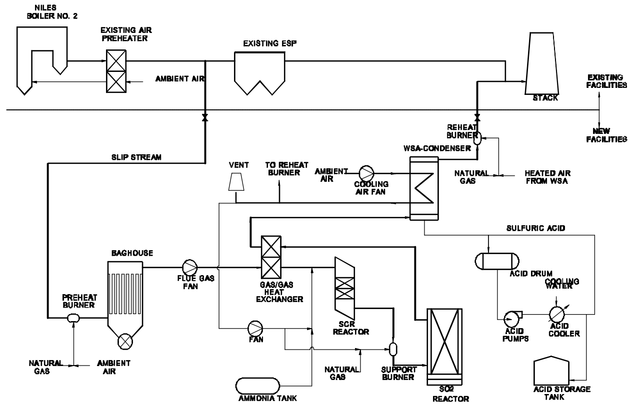


Also assisting the Project Manager was a Project Administrator who was responsible for overall administrative functions, including conformance of contract documents issuance of instructions to all participating departments, release of contract requisitions for material procurement, procurement monitoring, expediting and issuance of invoicing instructions.

A complete Project Organization is shown in Figure 1-1.

1.2.3 Project Description

The SNOX technology consists of four (4) key process areas which are NO_x reduction, SO_2 oxidation, sulfuric acid (H_2SO_4) condensation and acid conditioning. The integration of these individual steps is shown in Figure 1-2, which is the process flow diagram for the Niles Station.



In the SNOX process, the stack gas leaving the boiler is cleaned of fly ash in a high-efficiency fabric filter baghouse to minimize the cleaning frequency of the H_2SO_4 catalyst in the downstream SO_2 converter. The ash-free gas is reheated, and NO_x is reacted with small quantities of ammonia in a catalytic reactor, where the NO_x is converted to harmless nitrogen and water. The SO_2 is oxidized to SO_3 in a second catalytic converter. The gas then passes through a novel glass-tube condenser in which the SO_3 is hydrolyzed to concentrated H_2SO_4.

1.2.4 Site

For the SNOX Demonstration Project at the Niles Station, equipment and facilities are installed on a 150 ft by 120 ft previously unoccupied area southeast of the plant building. The Niles plant is situated on 130 acres along the southern bank of the Mahoning River and is part of the Ohio Edison System, which serves approximately 9,000 square miles in central and northeastern Ohio and western Pennsylvania (through its Pennsylvania Power Company subsidiary). Ohio is one of the leading states in the production and consumption of coal. In 1989, 31.4 million tons of coal were mined and reserves of 19 billion short tons have been estimated. Coal will likely continue to be the cornerstone of Ohio's energy supply. However, largely as a result of the Clean Air Act, the demand for Ohio-produced coal, both domestic and out-of-state, has fallen considerably since 1970.

One of eleven power plants in the Ohio Edison system, the Niles facility was commissioned in 1954. The main power plant structure covers an area of approximately 166 ft by 200 ft, and houses two cyclone coal-fired steam electricity-generating units with a net demonstrated total capacity of 216 MWe for both units. The boiler units burn high-sulfur coal with a capacity factor of approximately 67 %. Flue gases from both boiler units are dispersed into the atmosphere by a two-flue 393 ft tall stack. The plant utilizes two electrostatic precipitators to control particulate emissions. The SNOX Demonstration Project at Ohio Edison was designed to treat about one-third of the flue gas stream from Unit No. 2 or approximately 16 % of the total flue gas generated at the plant.

Niles is located in an industrialized section of Northeastern Ohio. The region has a large unemployed population caused by the decrease of heavy industry including steel manufacturers in the area surrounding Youngstown, Ohio. Skilled and unskilled labor for construction and operation of the SNOX system were used. Under-utilized skilled and unskilled labor was plentiful in the area. Utilities required for construction and operation of the plant are available at the site. The plant receives coal by truck and is thus readily compatible with virtually any coal source.

Niles Unit 2 is a coal-fired utility boiler firing eastern high-sulfur coal from Ohio and Pennsylvania. As a unit built before implementation of New Source Performance Standards (NSPS), it represents the class of boilers that is the primary target of DOE's program to develop clean coal technology. Cyclone fired boilers are concentrated in the Midwest, the region primarily suggested as being responsible for production of acid rain. These boilers account for only 8 % of the generating capacity of the United States, but produce about 18 % of the nitrogen oxides. In addition, these boilers fire high-sulfur, low fusion point Midwestern coal which can not be satisfactorily fired in other boilers. Sale of these coals would be restricted and coal mining in the Midwest would decrease if the coals are not burned in cyclone-fired boilers.

Another advantage of the site is its location in Ohio, which is one of the states considered to be a possible contributor to acidic precipitation in the Northeast and Canada. A demonstration site in Ohio was thus highly appropriate. Use of the Niles plant was supported by Ohio Edison, the State of Ohio, and the local coal industry, all of whom stand to benefit from a low cost method of suppressing acidic (NO_x and SO_x) emissions.

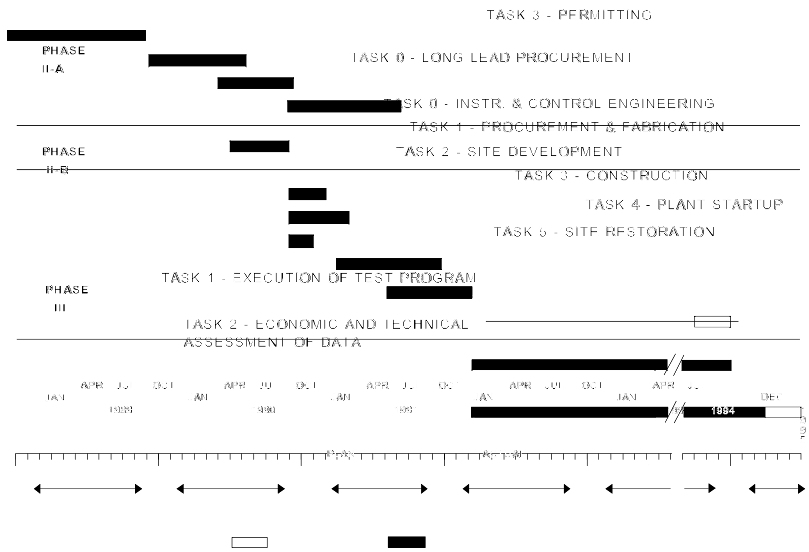
As mentioned above, Niles Unit 2 is a cyclone-fired boiler. Such boilers produce twice the level of NO_x emissions compared to pulverized coal-fired units. Methods of SO_2 or NO_x control that are available or being tested are difficult to apply to cyclone-fired boilers. Cyclone fired boilers reject most of the coal ash from the furnace bottom, and this design produces relatively low fly ash loading in the convective section and ESP. They are, therefore, not readily compatible with sorbent injection techniques for SO_2 control, which would represent almost an order of magnitude increase in solids throughput. Wet scrubbing of SO_2 could be added, but separate controls would be necessary for NO_x reduction, as opposed to SNOX which addresses both SO_2 and NO_x.

Reduction of NO_x emissions on Niles Unit 2 also benefits local air quality. Niles is in a non-attainment area for ozone, for which NO_x is a precursor. Reduction in NO_x emissions from a utility boiler is therefore of particular significance for air quality in this area.

1.2.5 Project Schedule

The execution of the SNOX Demonstration Project was divided into three phases which spanned approximately sixty months. These phases are identified as follows (shown in Figure 1-3):

- Phase I: Design and Permitting
- Phase IIA: Long Lead Procurement
- Phase IIB: Construction and Start-Up
- Phase III: Operation, Data Collection, and Reporting



Phase I of this project, Design and Permitting, was further broken down into Basic Engineering, Detailed Engineering and Permitting. Basic Engineering was completed in July of 1990, followed by the completion of the Detailed Engineering toward the end of that year.

Phase IIA was comprised of the procurement of long lead time items such as the baghouse, high temperature steel, control system, gas/gas heat exchanger, and the WSA condenser. These items were purchased at the beginning of Detailed Engineering and arrived at the Niles Plant for installation between February and May of 1991. Site preparation and installation of foundations began in November 1990 and construction was completed in November of 1991. Equipment commissioning was conducted following completion of construction and the system was first operated on flue gas in March of 1992.

Originally the overall program was 48 months in length and was scheduled to end in December of 1993. A task for Site Restoration was included in Phase II-B funding in the event that Ohio Edison did not opt to retain the plant. During the second half of 1993, Ohio Edison announced that it would retain the plant and funds that were designated for dismantling were reapportioned into the operating phase of the program for testing and system modifications. Part of Ohio Edison's decision to retain the plant hinged upon assurances by ABBES and DOE that existing auxiliary equipment and materials problems would be resolved. The test program was lengthened by twelve months to generate additional performance data and perform the system modifications.

Testing was completed through December of 1994, and the project is now complete.

1.3 Objectives of the Project

In order to demonstrate and evaluate the performance of the SNOX process during the CCT Program, operating data were collected and parametric tests conducted to characterize the process and equipment. The primary objectives for the SNOX Demonstration Project were as follows:

- Demonstrate NO_x and SO_2 removals of 90 and 95%, respectively
- Demonstrate the commercial quality of the product sulfuric acid.
- Satisfy all Environmental Monitoring Plan requirements.
- Perform a technical and economic characterization of the technology.

The following secondary objectives were identified in order to fully establish a basis for the technical and economic evaluation of a commercial application of this technology.

1. Execute parametric test batteries on all major pieces of equipment.
   - Fabric filter
   - SCR system
   - SO_2 converter
   - WSA condenser
   - Gas/gas heat exchanger
   - Catalyst screening unit
2. Quantify process consumptions.
   - Power
   - Natural gas
   - Catalysts
   - Cooling water
   - Potable water
   - Ammonia
3. Quantify process productions.
   - Sulfuric acid
   - Heat
4. Quantify personnel requirements.
5. Evaluate all materials of construction.

1.4 Significance of the Project

The SNOX Project was one of several CCT projects Ohio Edison currently has underway or has recently completed. Such experience with promising retrofit technologies that are simpler and less expensive than wet scrubbers will make acid rain compliance planning more efficient and effective for Ohio Edison.

The project served to demonstrate the high performance of the SNOX technology in the North American power generation industry, i.e. with U.S. fuels and operating staff. The first and foremost objective of the project was to successfully apply the technology and proprietary equipment to a power plant firing high-sulfur coal so as to confirm the capability to economically meet the pollution control needs of that market. Several supporting objectives were employed to meet this primary objective. These supporting objectives were: to confirm the results achieved at the Danish pilot facility, demonstrate the marketability and economic credits of the sulfuric acid and heat energy by-products, confirm the estimated low operating costs, and define any limitations of each piece of major equipment with respect to a utility environment.

Even though the technology had been fully characterized process-wise on pilot and prototype units prior to the contracting of the DOE project, the final scale-up to utility size required additional experience. Although all major components in earlier plants had been designed in a modular fashion utilizing full-scale components, scale-up would not be expected to result in any new process problems. However, achieving the correct distribution of flue gas or air through the various modules such as the WSA condenser, SO_2 Converter, and SCR remained to be demonstrated at full scale. This final scale-up experience was obtained through the DOE project.

The design of the DOE project was specifically aimed at demonstrating all SNOX subsystems and integrations that would be needed with a full-scale installation. For example, all flue gas conditions such as temperatures, pressures, and compositions were exactly replicated and all support systems, such as the control system, ammonia supply, and product acid storage and distribution were designed and operated as with a full-scale system. The only concept that could not be accommodated by the design was the integration of the WSA condenser discharge air as preheated combustion air for the boiler. This was not possible since the project was designed to treat only one third of the flue gas from the host boiler and would not result in a large enough quantity of combustion air. The principles involved with this integration, i.e. gas/gas heat exchanger design and preheated air for combustion, were believed to be fully understood and not crucial to the demonstration.

1.5 DOE's Role in the Project

The DOE was responsible for monitoring all aspects of the project and for granting or denying approvals required by this Cooperative Agreement. The DOE Contracting Officer was the authorized representative of the DOE for all matters related to the Cooperative Agreement.

The DOE Contracting Officer appointed a Contracting Officer's Technical Representative (COTR) who was the authorized representative for all technical matters and had the authority to issue "Technical Advice" which:

- Suggest redirection of the Cooperative Agreement effort, recommend a shifting of work emphasis between work areas or tasks, and suggest pursuit of certain lines of inquiry which assist in accomplishing the Statement of Work.
- Approve those technical reports, plans, and technical information required to be delivered by the Participant to the DOE under this Cooperative Agreement.

The DOE COTR did not have the authority to issue any technical advice which:

- Constitutes an assignment of additional work outside the Statement of Work.
- In any manner causes an increase or decrease in the total estimated cost, or the time required for performance of the Cooperative Agreement.
- Changes any of the terms, conditions, or specifications of the Cooperative Agreement.
- Interferes with the Participant's right to perform the terms and conditions of the Cooperative Agreement.

2.0 TECHNOLOGY DESCRIPTION

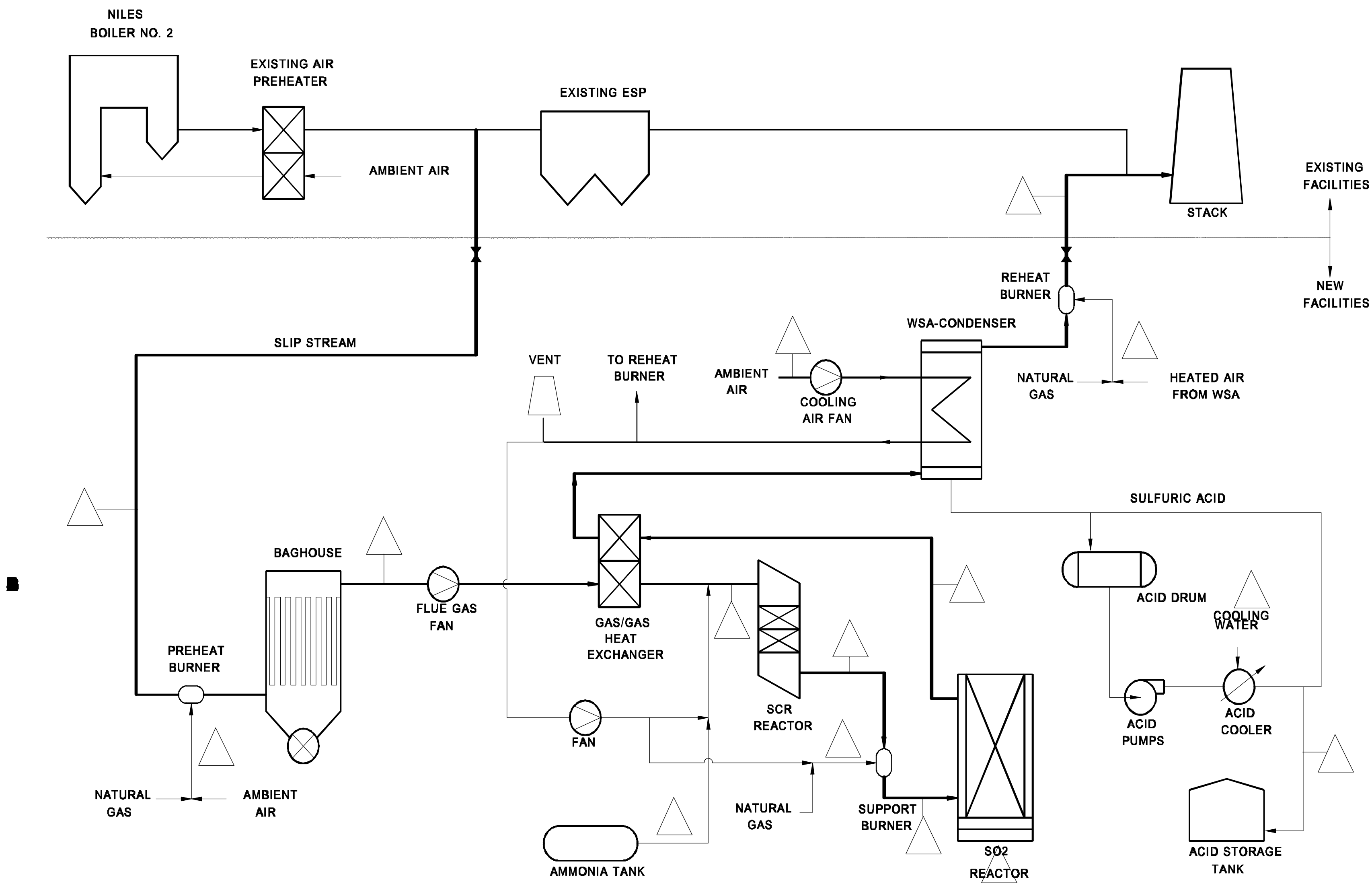
2.1 Description of the Demonstrated Technology

The SNOX technology consists of four (4) key process areas which are NOx reduction, SO_2 oxidation, H_2SO_4 condensation and acid conditioning. The process flow diagram for the Niles Station is shown in Figure 2-1, with stream flows and compositions in Table 2-1.

To follow Figure 2-1, a slip stream of flue gas which leaves the air preheater prior to the existing ESP is ducted to a preheat burner and fabric filter. The preheat burner raises the temperature of the flue gas to about 400EF (204EC) to simulate a full size, integrated system wherein preheated air from the SNOX plant would be fed into the boiler's air heater and would raise the outlet flue gas temperature. (The hot air at the demonstration plant is vented to atmosphere.)

After exiting the baghouse the flue gas is passed through the primary side of a gas/gas heat exchanger (GGH) which raises the gas temperature to above 700EF (370EC). A mixture of ammonia (NH_3) and air is added to the gas prior to the selective catalytic reduction (SCR) reactor, where nitrogen oxides are reduced to free nitrogen and water. After the flue gas leaves the SCR, its temperature is adjusted slightly by the second support burner, and it then enters the SO_2 converter which oxidizes SO_2 to sulfur trioxide (SO_3). The SO_3 laden gas is passed through the secondary side of the GGH, where it is cooled as the incoming flue gas is heated.

The processed flue gas is then passed through a falling film condenser (the WSA condenser) where it is further cooled with ambient air to below the sulfuric acid dewpoint. Acid condenses out of the gas phase on the inside of the glass tubed condenser and is subsequently collected, cooled, diluted, and stored. Cooling air leaves the WSA condenser at over 400EF (200EC). At the Niles plant, most of it is vented although a small amount is used for combustion air to the natural gas burners and for dilution air for the ammonia injection into the SCR. As mentioned above, in a full-size, integrated installation, the heated air from the condenser would be used as combustion air to the boiler.


NILES
BOILER NO. 2
EXISTING AIR
PREHEATER
EXISTING ESP
AMBIENT AIR
STACK
EXISTING
FACILITIES
NEW
FACILITIES
REHEAT
BURNER
WSA-CONDENSER
SLIP STREAM
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ACID DRUM
COOLING
WATER
PREHEAT
BURNER
SCR
REACTOR
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COOLER
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NATURAL
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SO2
REACTOR
ACID STORAGE
TANK
AMMONIA TANK

Table 2-1 Process Stream Flows and Compositions - Projected Normal Operating Conditions

Loc.	Pres.	Temp	Stream Flowrate - lbs per hour											
	in w.g.	E F	O_2	H_2O	CO_2	SO_2	SO_3	NH_3	H_2SO_4	NOx	N_2	CH_4	Ash	Total
1	0	270	18530	15202	71664	1923	24			444	251224		50.000	359061
2	6	400	18682	16070	72715	1933	24			444	256612		0.300	366480
3	10	730	19512	16098	72715	1933	24	167		444	259358		0.300	370251
4	16	733	19436	16352	72715	1933	24	8		22	259614		0.300	370104
5	17	780	19547	17021	73529	1933	24	8		22	263759		0.300	375843
6	23	785	19076	17032	73529	91	2315			36	263763		0.030	375842
7	33	212	19846	17032	74621	91	12			36	271396		0.003	383034
8	atm	60	76579	2519	153						252136			331387
A	5	60	1637	62							5388	382		7469
B	5	60	1259	39							4145	296		5739
C	5	60	1701	64							5600	397		7762
D	40 psi	60						167						
E	50 psi													
F	20 psi			213					2832					3045
G	atm												0.297	0.297

2.2 Detailed Process Information

2.2.1 Particulate Collection

The degree of particulate collection upstream of the reactor has a significant effect on the operating costs of this process. This correlation is due to the inherent characteristic of the SO_2 converter catalyst to collect and retain greater than 90% of all particulate matter which enters the converter. The collection of this particulate matter, over time, increases the pressure drop across the SO_2 converter. The pressure drop can, however, be restored through catalyst screening (described later). Higher dust loads therefore require more frequent catalyst screening which provides an incentive to utilize a high efficiency particulate collector upstream of the SNOX process area. A dust level of 1 mg/Nm^3 leaving the collector has been targeted. A fabric filter with PTFE membrane bags has been demonstrated to achieve this very low emission level. As a consequence of both the high efficiency dust collector and the dust retention characteristics of the SO_2 converter, particulate emissions from the system were predicted to be significantly less than 1 mg/Nm^3 (0.0004 gr/SCF), which is far below any current regulations or standards. The process and instrumentation diagram (P&ID) for this section of the process is shown in Figure 2-2.

It should be noted that while a high efficiency particulate collector has benefits related to system operating costs, the economics do not require it exclusively. The Niles Demonstration Project with a fabric filter and PTFE bags was expected to require screening only once per year. Low dust levels entering the catalyst areas have an additional benefit of eliminating the capital and operating costs associated with steam cleaning lances that may be necessary for the GGH and SCR at higher dust levels.

Figure 2-2 Flue Gas Dedusting (P&ID)

2.2.2 Nitrogen Oxide Reduction

After the particulate matter is collected and the temperature of the flue gas is increased to over 700EF (370EC) through the GGH, an ammonia and air mixture is introduced to the gas stream through a proprietary nozzle grid located upstream of the SCR. The proprietary design of the nozzle grid allows for controlled stoichiometric ratios of NH_3 to NO_x on a localized scale over the cross-section of the SCR inlet duct. This is critical in order to optimize system NO_x removal efficiency. Further, any unreacted NH_3 which "slips" across the SCR will be oxidized to NO_x, water, and N_2 in the SO_2 converter downstream. A slipstream of hot air which leaves the WSA condenser is used to evaporate and dilute a metered amount of ammonia. This NH_3/air mixture, which must be below 12 vol.% NH_3 to avoid an explosion hazard, is agitated with a static mixer and supplied to the distribution grid.

The flue gas/NH_3 mixture enters the SCR unit and contacts the Haldor Topsoe DNX monolithic catalyst, which has been demonstrated to remove up to 97%+ of the entering NO_x. The reduction follows Equation 2-1.

Install Equation Editor and double-click here to view equation.

In this equation, NO is taken to represent NO_x. The small amount of NO_2 present in the flue gas is reduced similarly.

The general arrangement of the SNOX process offers a significant advantage over other SCR technologies using NH_3 in that those processes are limited to NH_3/NO_x molar ratios of less than 1.0. This must be done in order to limit the NH_3 "slip" past the SCR to 5 ppm or less. Higher levels may result in ammonium bisulfate or bisulfite scaling, which can become a problem at high No_x removal efficiencies, typically 90 or greater. Any NH_3 slip in the SNOX process, however, is oxidized as it contacts the SO_2 converter catalyst downstream. This allows stoichiometric ratios of 1.00 to 1.05 and consequently higher NO_x removal efficiencies without the adverse downstream effects of higher ammonia concentrations. Ammonium "salting" in the duct between the SCR and the SO_2 converter in the SNOX process is avoided by the high temperature in this area. Excess NH_3 slippage, however, must still be minimized in order to maximize system No_x removal due to partial oxidation of NH_3 to No_x in the SO_2 converter downstream. The P&ID for the SCR reactor is shown in Figure 2-3, and the ammonia storage P&ID is in Figure 2-4.

2.2.3 Sulfur Dioxide Oxidation

The SCR effluent is again heated slightly with natural gas, oil or steam to reach the optimum SO_2 converter inlet temperature and passed through Haldor Topsoe VK WSA sulfuric acid catalyst. The VK WSA catalyst belongs to Topsoe's VK sulfuric acid catalyst series which has seen wide use in the U.S. sulfuric acid industry for the past decade with a high degree of success. Without any reagents or additives, over 95% of the entering SO_2 is oxidized via Equation 2-2.

Install Equation Editor and double-click here to view equation.

The efficiency of the Topsoe catalyst is not affected by the presence of water vapor or chlorides in concentrations up to 50% and several hundred ppm, respectively.

Due to surface fouling by fly ash, the SO_2 converter catalyst requires screening to maintain oxidation efficiency at a frequency dependent on the removal efficiency of the particulate collection device upstream. The required screening frequency will range from once every two weeks to once a year. Regardless of the efficiency of the particulate collector, however, virtually all remaining particulate matter is retained in the SO_2 reactor, which results in the inherently minimal particulate emissions of this process but also necessitates catalyst screening on some regular basis.

Figure 2-3 NO_x Reduction (P&ID)

Figure 2-4 Ammonia Storage Section (P&ID)

The screening procedure consists of isolating an individual catalyst bed, removal and mechanical screening of the catalyst in that bed, and refilling the bed with the screened catalyst. Beds are sequentially screened in this manner. The procedure can be automated and is performed while the process is on-line. This procedure separates and removes virtually all fly ash and other contaminants from the surface of the catalyst pellets and restores the pressure drop across the SO_2 converter. Catalyst loss during screening is estimated at 2-3%.

An additional benefit of the sulfuric acid catalyst is its ability to oxidize carbon monoxide (CO) and hydrocarbons present in the flue gas stream to carbon dioxide and water. This is of importance since hydrocarbons and CO emissions have come under increased scrutiny, as evidenced by passage of the Clean Air Act Amendments. The P&ID for the SO_2 conversion section of the process is shown in Figure 2-5.

2.2.4 Sulfuric Acid Condensation

The SO_3 in the gas leaving the SO_2 converter is hydrated and condensed in two steps. First, the bulk of the SO_3 is hydrated to sulfuric acid vapor (Equation 2-3) as the flue gas passes through the secondary side of the GGH and drops approximately 300EF (170EC).

Install Equation Editor and double-click here to view equation.

At this point the flue gas is still well above the acid dewpoint, thus avoiding acid condensation and corrosion of the ductwork. The flue gas is then passed through the proprietary WSA condenser developed by Haldor Topsoe where the gaseous H_2SO_4 is converted to liquid H_2SO_4 according to Equation 2-4. The WSA condenser is a unique tube and shell falling film condenser with ambient air used as a cooling medium on the shell side. Proprietary borosilicate glass tubes are used to

Install Equation Editor and double-click here to view equation.

convey and cool the flue gas. There are several features of these tubes which make possible the

Figure 2-5 SO_2 Conversion (P&ID)

virtually complete condensation and capture of the sulfuric acid at concentrations of 94 to 97 wt.%. The flue gas is cooled to about 212EF (100EC) at the outlet of the condenser. This, combined with the presence of approximately 5ppm of uncollected sulfuric acid mist, will require the downstream ductwork and stack to be lined. The condensed sulfuric acid product is funneled through an acid brick lined trough at the bottom of the WSA condenser into the acid conditioning and storage system.

The WSA condenser's discharge cooling air represents the only other by-product of the SNOX process. In an integrated system, the bulk of this heated ambient air at about 400EF (200EC) will be passed through the furnace air preheater and used as combustion air. A small percentage of the product air is used for system auxiliaries such as ammonia evaporation and dilution, natural gas burner combustion air, and catalyst screening equipment heatup. The WSA condenser, in effect, collects the heat released from the reactions in the SCR and SO_2 converter, the hydration of SO_3, the condensation of H_2SO_4, the support burner, the booster fan compression and the overall decrease in flue gas temperature. This results in a considerable amount of energy which can be easily utilized as preheated combustion air in the furnace to increase boiler heat rate. This, in turn, increases overall thermal output per unit of fuel.

The P&ID for the acid condensation section is shown in Figure 2-6.

2.2.5 Acid Conditioning and Storage Systems

The hot concentrated H_2SO_4 product at about 400EF (200EC) is collected and circulated through a thermoplastic lined system consisting of a holding tank, circulation pumps, and a water cooled tube and shell heat exchanger. The purpose of this loop is to cool the acid to more conveniently manageable temperatures (70-100EF, or 20-40EC) and to allow for dilution of the acid to the commercially traded concentration of 93.2 wt.%. As acid is collected in the conditioning loop, it is metered off at a rate which maintains the level in the holding tank. This product acid is stored in lined carbon steel tanks prior to removal by tanker truck. Based on prototype results, the sulfuric acid produced by the SNOX process at the Niles station is expected to meet or exceed

Figure 2-6 Acid Condensation (P&ID)

U.S. Federal Specification O-S-80C Class 1 and be commercially tradeable without limitation. The P&ID for this section is shown in Figure 2-7.

2.2.6 Heat Addition, Transfer, and Recovery

Heat addition, transfer and recovery are of significant concern in the SNOX process as they influence the economic success of the technology. The process requires heat only to trim the flue gas temperature between the SCR and the SO_2 converter in a full size, integrated installation. It is anticipated that the most efficient and cost effective source of this heat in a utility environment will be steam, however, natural gas or oil can be effectively utilized.

One major piece of equipment in the SNOX process is the GGH, which allows the use of the high temperatures in the process area in an economic manner by transferring sensible heat in the treated flue gas stream to the process inlet stream. Without the recycling of this energy, it is doubtful that the process would be economically viable. Selection of the type of heat exchanger is of concern since any leakage of flue gas across the GGH, as with a rotary-type, would bypass both reactors and result in lower measured system removal efficiencies. This effect would be proportional to the amount of leakage in these heat exchangers. A stationary-type or heat pipe GGH, however, has been researched and determined to be a superior technology for the demonstration project due to its inherent zero leakage between the two gas paths and lack of moving parts, allowing maximum system performance.

The process generates recoverable heat in several ways. The two reactions via Equations (2-1) and (2-2) increase the flue gas temperature by about 22EF (12EC) per 1000 ppm of NO_x and by about 5EF (3EC) per 1000 ppm of SO_2. Energy is further released by the hydration (Equation 2-3) and condensation (Equation 2-4) of sulfuric acid. In total, sulfuric acid production yields 2.42 kWh(th) per kg (3,750 Btu/lb) of sulfur recovered as acid. The above heats plus the heat of compression generated by the flue gas booster fan and the support heat added after the SCR are recovered in the WSA condenser cooling air discharge for use in the furnace as combustion air.

Figure 2-7 Acid Cooling and Storage (P&ID)

This can increase steam production on the order of 1% per each percent of sulfur in the fuel. At 2-3% sulfur, the recovered heat is equivalent to the SNOX process energy requirements.

The above characteristics make the use of high-sulfur, low cost fuels economically attractive with this process. It should be noted that in a retrofit application of this technology, some modifications to the existing power block equipment will be required in order to fully benefit from the new combustion air source. First, modification of the furnace airpreheater may be necessary due to the increased volume and temperature of the inlet air. Second, the increased boiler heat rate may require additional tube banks in the boiler if the design fuel feed rate is to be maintained. Of course, the thermal output of the boiler will increase accordingly.

3.0 UPDATE OF THE PUBLIC DESIGN REPORT

3.1 Design and EquipmentChanges

The SNOX system design, as set forth in Final Report Volume I: Public Design, remains valid for this report. However, some equipment and material failures occurred that necessitated repairs and/or replacement. The following discusses these equipment changes.

Ammonia Pump - replacement

Problems persisted with the originally installed ammonia pump throughout the initial operating period of the project. Primarily, these problems revolved around the pump losing prime, internal flashing, and subsequently failing to maintain adequate ammonia flow to the system. The original diaphragm type pump was replaced with a geared rotary design which proved far superior in operation.

Burners - replacement

Burners No. 1 and 3 experienced numerous failures throughout the initial operating period. These failures included:

> <u>Constant tripping</u> -- During operation, the burners would intermittently trip off line, requiring manual operator intervention to re-start. Various fixes were tried, including changes to and replacement of the flame detector circuits.
>
> <u>Failure to reach operating temperature</u> -- On numerous occasions, especially during the start-up phase of the project, the burners failed (were very slow) to reach proper operating temperature. These failures proved to be critical to the operation and maintenance of the system as can be seen in the discussions below concerning the PTFE bag deterioration in the fabric filter and acid deterioration of the ductwork lining. Again, various fixes were

tried including the design and installation of cone shaped shrouds on the burners to attempt to modify the flame pattern. These fixes helped but created problems themselves by causing flame impingement on the ductwork. The burners, and associated controls, were eventually replaced with a different design.

Fabric Filter - PTFE bag replacement

During operations, after initial start-up, excessive differential pressure readings were noted across the fabric filter. Additionally, particulate measurement at the fabric filter outlet exceeded that which was expected based on the design parameters and the operating characteristics of the PTFE bags. Neither of these aforementioned problems were of serious concern and did not affect the overall operation of the system or affect the gathering of test data. Therefore, it was decided to continue operations and inspect the bags during a future planned outage.

Upon inspection, it was determined that the bags had deteriorated to the point that small perforations were present and major failures might occur in the near future. Analysis led to the conclusion that improper operating temperatures during the initial start-up of the system (see burners above) caused acid rich particulate caking, which led to general bag deterioration. The bags were replaced in total, and the fabric filter operated without further problems until the conclusion of the project.

Flue Gas Outlet Duct - Relining, Replacement

Replacement - The flue gas ductwork lining between the WSA condenser and the No. 3 burner was found to be excessively deteriorated. The failure of the lining caused corrosion of the carbon steel duct. New ductwork was fabricated and installed. Although the duct metal was not perforated, it was determined that it was less costly to have new ductwork fabricated than to clean and recoat the old ductwork in place; this would have also risked subsequent coating failure due to imperfect surface cleaning/preparation.

Relining - It was determined that the original lining was unable to withstand the acid concentration of droplets in the outlet ductwork. Although the amount of acid carryover is very small, the walls of the duct do become wetted and the acid is too concentrated (approx. 70%) for some lining materials. After investigation of various linings, a vinyl ester coating was chosen, a composition which had ultimately proved to be satisfactory at the Danish SNOX plant. The new ductwork was fabricated, coated, and installed.

Expansion Joints - Replacement

Problems occurred with the duct expansion joints from the onset of operations. After operating for only a few weeks, leaks were discovered at various expansion joints in the system.

The first attempts at repairs included simply tightening fasteners and installing additional acid drains at the expansion joints. This however proved inadequate, and replacement of the joints commenced, first with joints of like material and design, then with various gasket materials, etc. Each attempt was the result of consultations with numerous expansion joint manufactures with considerable experience in the field. Expansion joint problems were resolved at all locations except the outlets of the SO_2 converter. This location is particularly subject to aggressive corrosion because of the high temperature (780E F) and high SO_3 concentrations.

New expansion joints, of the seal air type versus baffle and pillow, were ordered and installed. This type of expansion joint had proven effective at the Danish SNOX plant where similar problems had occurred. Concurrent installation of extensive seal air piping to serve this type of expansion joint was also required. After completion of installation, operations resumed. Thus far, the new expansion joints have proven satisfactory, with no leakage occurring from installation to the end of the test program.

WSA Tower - inlet nozzle repair

An acid leak that developed at the condenser flue gas inlet was investigated. Upon removal of the associated lagging and insulation, it was found that the metal bottom of the inlet nozzle was corroded through due to flue gas leaking past the grout joint between the acid brick and metal flange.

Large diameter PTFE gaskets were installed in all four of the WSA condenser inlets to seal the space between the acid brick and the metal at the ends of the inlets. Additionally, new grout was installed at the ends of the condenser and a reinforcing section of semi-circular metal sheets was installed over the original metal housing of the condenser inlets.

Flue Gas Booster Fan - bearing failure

Intermittent problems occurred with the bearings on the flue gas booster fan during the course of the program. These problems primarily concerned bearing temperature and subsequent failure.

It was speculated that the oiling system in the original design was inadequate for the loads placed on the fan. A force fed oil system was installed, plus new bearings, which appear to have solved the problem.

Acid Holding Tank - lining

Inspection of the acid holding tank revealed that the tank lining was deteriorating and required replacement. A stronger, less pregnable lining material was selected and installed.

Additional Electrical Modifications and Installation

Related to the above installations, considerable electrical work was performed. This involved the installation of heaters, blower wiring, heat tracing, and sensors and associated wiring for the seal air system, and wiring for the force fed oil system.

Additionally, throughout the program, modifications/additions/deletions of test sensors were required. The changes made to the test sensors however, were related to the test program and not the design and operation of the system.

3.2 Demonstration Plant Capital Cost Update

The costs associated with the repairs and modifications noted in Section 3.1 above are detailed in Table 3-1. Cost of the initially installed SNOX plant was $20,011,004, as detailed in the Public Design Report, including engineering but without general and administrative expenses (G&A). The additional capital cost items described above total $846,850 and bring the final installed plant cost of the SNOX system to $20,858,000.

3.3 Demonstration Plant Operating Costs Update

No significant changes to the operating costs as projected in the Public Design Report occurred during the operation of the plant.

Table 3-1 SNOX Plant Additional Capital Cost

MODIFICATION	COST (MATERIAL & LABOR)
Ammonia Pump	$1,060
Burners	$57,680
Fabric Filter Bags	$99,490
Duct (incl. lining & insulation)	$316,700
Expansion Joints (incl. seal air system)	$276,620
WSA Tower	$10,130
Flue Gas Booster Fan	$24,000
Acid Holding Tank	$7830
Electrical	$53,340
Total Additional Cost	**$846,850**
SNOX Plant Initial Installed Cost	$20,011,004
SNOX Plant Final Installed Cost	**$20,857,854**

4.0 DEMONSTRATION PROGRAM

This section discusses Phase III, Operation and Testing, that was implemented as part of the demonstration project.

The primary objective of the overall test plan developed and executed during the demonstration project was to demonstrate and evaluate the performance of the SNOX process in a North American high-sulfur, coal-fired commercial application.

Specifically, the test plan was designed to:

- Demonstrate NO_x and SO_2 removals of 90 and 95%, respectively.
- Demonstrate the commercial quality of the product sulfuric acid.
- Perform a technical and economical characterization of the technology.

This section primarily discusses the testing accomplished to satisfy the first two requirements plus the technical characterization of the SNOX process elements. The economic characterization is discussed in Section 7.0.

4.1 Test Plans

The Data Collection Plan, which includes specific details of the test batteries conducted during the demonstration project, can be found in Appendix A. A review of the tests conducted will reveal that numerous parameters were varied throughout the program to effectively demonstrate the viability of the process and design of the plant.

Parametric test batteries were conducted on the following pieces of major equipment:

- Fabric filter
- SCR system
- SO_2 converter
- WSA condenser
- Gas/gas heat exchanger

! Catalyst screening unit

Process temperatures and flow rates were varied during the test program. These tests were conducted to determine the performance of the process under various conditions. It was important, for example, to determine the effect of different ammonia injection stoichiometries on the performance of the system as related to NO_x reduction and ammonia slip. Temperature variations were used likewise to determine catalyst efficiency effects for both the SO_2 and NO_x reactors.

In many of the tests conducted, the primary parameter varied was the actual flue gas load admitted to the plant. This was an important variable for many of the process areas. For the NO_x and SO_2 reactors, changes in gas flow affect the space velocity in the catalyst beds. In the WSA condenser, total system gas flow affects the linear flue gas velocity within the glass condenser tubes, thereby influencing heat transfer and droplet collection efficiency. Load was varied from approximately 50 to 110% of SNOX plant design capacity.

Table 4-1 lists the parameters varied and the range of variance introduced during testing.

4.1.1 Short Term Tests

Table 4-2 Short Term Tests

TEST NAME	VARIABLE	RANGE
Baseline Testing	None	Unit 2 at full load
Fabric Filter Parametric Study	Design load	75%, 100%, 110%
SCR System Parametric Study	NH_3/NO_x molar ratio	0.95 - 1.10
SO_2 Converter Parametric Study	Design load Inlet temperature	80% - 110% 715E - 800E F
WSA Condenser Parametric Study	Design load Flue gas outlet temp.	60% - 110% 200E - 230E F
Optimized System Tests	Design load	30% - 100%

The above Table 4-2 lists the short term testing accomplished on the Demonstration Plant. Short term testing was used to characterize the major components of the plant and to construct a base line from which to compare longer term, overall operating characteristics. The tests, taken in the order presented, had the following objectives:

Baseline Testing -- encompassed manual testing for the characterization of the existing Unit 2 process streams, calibration and verification of all instruments, and initial evaluation of the as-installed performance of all equipment. Baseline testing was broken down into three test groups:

2A -- Unit 2 characterization of the existing Unit 2 ESP inlet and stack flue gas streams.
2B -- Calibrations of the baghouse outlet dust monitor, venturi, continuous emissions monitors (CEMS) and Acid concentration monitor
2C -- General equipment performance to evaluate the as-installed performance of each major piece of equipment.

Fabric Filter Parametric Study -- was conducted to characterize the performance of the fabric filter. The filter was tested for particulate collection efficiencies at three load points, thereby varying the air-to-cloth ratio of the filter. The efficiency of the fabric filter directly impacts catalyst screening frequency and life expectancy.

SCR System Parametric Study -- was executed to characterize the SCR reactor, which includes the SCR vessel itself and the ammonia injection apparatus. This test was divided into four main groups which were identical with the exception that they were executed periodically over the test program to quantify any catalyst degradation. Ammonia stoichiometry was varied and NO_x reduction efficiency was measured along with ammonia slip.

SO_2 Converter Parametric Study -- was accomplished to characterize the SO_2 oxidation efficiency of the SO_2 converter as a function of catalyst bed operating temperature and flue gas throughput (space velocity).

WSA condenser Parametric Study -- was used to characterize the performance of the WSA condenser with respect to SO_3 condensation efficiency, H_2SO_4 mist carryover, and product acid quality. These tests were conducted as a function of flue gas flow (design load) and flue gas outlet temperature. Flue gas flow affects the heat transfer within the glass condensing tubes as well as potential carryover. Flue gas outlet temperature affects vapor pressure and therefore concentration of gaseous H_2SO_4/SO_3 in the flue gas. Additionally, the effects of cooling air temperature on condenser performance (primarily acid mist formation) were studied. This portion of the test was run separately from the aforementioned characterization study as winter operating conditions were required.

The complete test parameters for the summarized tests can be found in Appendix A. The variables and plant conditions chosen for the individual test matrixes were designed to most closely mimic the operating conditions likely to be encountered during a commercial plant's operating cycle. Even though the SNOX Demonstration Plant was designed to operate on one third of the available flue gas of the Niles Unit 2 boiler, the components were full-scale, commercial units.

TEST NAME	PARAMETER	RANGE
System Performance Analysis	Design load	75%, 100%, 110%
Optimized System Tests	Design load	30% - 100%

4.1.2 Long-term Tests

The tests identified in the Table 4-3, along with the continuous recording of the data at the distributive control system (DCS), were utilized to determine the stability of the process and the long term operability. In addition to these data, periodic inspection of the mechanical components, including ductwork, piping and expansion joints, was made; test coupons removed for analysis; and routine analysis of the acid produced for clarity and strength was accomplished.

Load following characteristics of the SNOX plant were also investigated during these tests. Automated load following capability was included in the DCS. Another purpose of extended run time was to evaluate any degradation in both the SCR and SO_2 catalysts.

4.2 Operating Procedures

The following discussion is a general outline of the start-up, operation, and shutdown procedures for the SNOX Demonstration Plant and selected plant sub-systems. With minor exceptions, operation of the plant is automatic and in fact only a single operator is actually required to run the plant.

A. Start-Up

1. Acid System

The acid circulation and cooling system must always be filled with acid and the pumps running before any flow through the flue gas system is established. Cooling water flow to the acid cooler must be established also.

2. Burners

Before flue gas can enter the SNOX plant, the three burners (one support and two reheat) must be placed in operation while the plant is brought up to operating temperature on air. The burners ensure that operating temperatures necessary for the process reactions to occur are established, and that unwanted condensation of water vapor or SO_3 does not occur when flue gas is introduced into the system.

3. Fans

The cooling air fan, ammonia mix blower, and flue gas booster fan are started. After the fans have been started, an ambient air flow of approximately 3000 lbs./min is established.

4. Attainment of Operating Temperatures.

Admission of flue gas to the SNOX plant can take place only if equipment, piping and catalyst have been preheated to predetermined temperatures. As ambient air is flowing through the SNOX plant, the equipment is heated by the natural gas fired burners (see 2 and 3 above), which are ramped at a rate to prevent damage to the condenser brick lining and other components. Preheating the plant allows the catalysts to attain reaction temperatures and prevents unwanted water or acid condensation in the gas/gas heat exchanger and other components and ductwork when flue gas is introduced.

Three primary conditions must be met before the plant may be switched to flue gas:

! SCR outlet temperature should exceed 500EF.
! SO_2 converter outlet temperature should exceed 585EF.
! The gas/gas heat exchanger skin temperature should exceed 400EF.

5. Admission of Flue Gas

Once the plant has reached operating temperatures and stabilized, flue gas may be admitted to the system. This is accomplished gradually, with a simultaneous reduction in the flow of ambient air. During this phase of the start-up sequence, temperatures and other process data are closely watched, and adjustments made accordingly. As the plant reaches 100% operation on flue gas, fully automatic operation can be established.

B. Operations

As noted above, operation of the plant normally requires little manual adjustment. The operator functions primarily as a monitor to ensure that the system remains within operating parameters.

Manual operation may be required if problems occur with system sensors and automatic control becomes problematic.

All critical components and process areas are remotely sensed and alarmed through the DCS. This includes critical temperatures and pressures throughout the system, both process related and mechanical. In addition, the operator has available readouts from the CEM system. Readings from this system gives the operator the ability to access the overall operating efficiency of the system.

Additionally, the operator makes routine walkdowns of the system area, using a checklist to inspect for visual and aural abnormalities. With the information made available to the operator he/she can quickly react to an anomaly by either manually intervening and adjusting the system, or in the case of a critical failure, performing a shutdown.

C. ShutDown

Shutdown is essentially the reverse of the start-up procedure. The plant is switched to ambient air and the burners ramped to allow cooling of the SO_2 converter to a temperature below which desorption of SO_3 is negligible and it is no longer necessary to keep the converter gas side duct work above the sulfuric acid dew point. If the shutdown is for a short duration, temperatures can be maintained with the support burners while on air to allow rapid reintroduction of flue gas when desired.

4.2.1 Instrumentation and Data Acquisition

Listed below, with a detailed narrative explanation, are the major control and instrumentation areas of the SNOX Demonstration Plant. When viewed in total, it can be seen that the SNOX

Demonstration Plant is virtually automatic, requiring operator intervention only when equipment or operating anomalies occur. Due to the alarm and information display system incorporated into the DCS, a vast majority of operator required interventions can be handled directly in the control room.

Only minor changes from the original design and installation were made during the Demonstration Project. The most significant change was to the burner controls, which were replaced because of reliability problems with the original equipment. The stability of the process as a whole, after the initial operating period, was excellent, with all process controls functioning as designed.

Maintenance required of the control system was routine. Occasional blockages of flow or pressure metering points was experienced but no more than would be expected. Additionally there were minor problems with control linkages binding, or adjustment on the various dampers. Significantly, problems in this area were immediately evident by indicators or readings available and monitored in the control room, allowing rapid response and repair to occur.

The single change of significance, recommended for follow-on commercial installations, is to the CEM system. As designed, the CEM system for the SNOX Demonstration Plant is in a centralized location and used conventional extractive sampling methods without dilution. The analyzer location at the Demonstration Plant requires long runs of sample tubing to the various test locations. These long runs have proven to be prone to clogging during normal operations. For future installations, the analyzers should be located closer to sampling points, a dilution type system employed, or an in-situ measurement probe employed.

Major plant instrumentation and control areas are discussed in the following sections:

<u>Dampers (Inlet ZCV-0100, Inlet Vent ZCV-0120, Outlet ZCV-0108, Outlet Vent ZCV-012</u>1

The dampers are adjustable throughout the open-closed range and are controlled by the DCS through forward/reversing motor starters in the motor control center and AUMA actuators at the dampers. The DCS receives continuous feedback signals representing 0-100% open. Limit switches are hardwired to the motor starter for equipment protection. Seal air is locally indicated with pressure gauges and low pressure switches alarm the DCS for low seal air pressure.

Support Burners (No. 1 H-201, No. 2 H-207, No. 3 H-210) The SNOX plants three support burners have basically identical controls and control instrumentation:

- Burner inlet temperatures and pressures are measured and recorded by the DCS.
- Outlet temperature is sensed and recorded by the DCS
- The DCS provides the local burner control panel with a signal corresponding to burner load.
- The local burner control panel controls the fuel and combustion air flows to achieve the desired output and provides local safety protection.
- The burners are capable of automatic control by the DCS or manual control.
- The burners are started through the DCS.
- The local control panel relays back a signal representing "Burner Flame on/Off"
- The outlet temperature is the feedback control variable to the DCS for modulating the burner's heat input.

Baghouse (A-202) The baghouse is controlled independently by the baghouse control panel, located in the control room. The three principal areas of control, accomplished by a General Electric sequencing logic controller, are module isolation, cleaning, and hopper heating. The DCS also receives the baghouse differential pressure from the control panel. This signal represents 0-10" w.c. of pressure drop and the cleaning cycle maintains a set point of 5"w.c. An Auburn International Broken Bag Detector is mounted in the outlet duct of the baghouse which transmits a signal corresponding to the number of grains of flyash per cubic foot of flue gas. The 1st support burner outlet temperature element and the venturi inlet temperature elements transmit baghouse inlet and outlet temperatures to the DCS. The calculated difference in temperature is used to detect any fires in the baghouse.

NH_3/Air Mixing Unit (X-240) The flow of air is measured by the mixing air flow element and transmitted to the DCS. This signal corresponds to 0-1000 acfm. The flow of air is controlled through the mixing air flow control valve with a set point determined by the mass flow of ammonia and the maximum NH_3 concentration of 6% (vol). After the mixture exits the ammonia/air mixing unit, the mixture temperature is measured by the ammonia/air temperature

Element and recorded by the DCS. Over-pressure protection of the mixture is provided by a pressure safety valve (PSV-0508).

Selective Catalytic Reduction Reactor (R-206) The SCR Reactor is monitored for both pressure drop and inlet and outlet temperature. An expected temperature increase of 14EF will occur across the SCR reactor due to the NO_x reduction reactions. The inlet and outlet temperatures are recorded by the DCS. The differential pressure across the reactor is monitored by the differential pressure transmitter.

Ammonia Pump (P-230) The pump outlet pressure is regulated by the ammonia feed pump outlet pressure regulator. The ammonia feed pump Outlet Pressure gauge indicates the ammonia supply pressure. The mass flow of ammonia is sensed by the Fischer & Porter Coriolis flow element and transmitted to the DCS by the ammonia flow transmitter as a signal representing 0-250 pounds per hour. The mass flow of ammonia is controlled by the DCS and is calculated from: NO_x signals from the gas analyzer, flue gas mass flow signal from the Venturi, boiler load from Ohio Edison, and NH_3/NO_x molar ratio set point. The ammonia atomizer is a Yarway De-superheater/Attemporator which atomizes the liquid ammonia and controls the amount of ammonia injection based on the above calculation.

SO_2 Converter (R-208) The SO_2 converter inlet and differential pressures are monitored and transmitted to the DCS. Thermowells are provided in each of the five inlet ducts for test measurements. Similarly, each of the 8 beds of the converter are equipped with two thermowells for test measurements. The SO_2 Converter has four outlet ducts. The temperature of each duct is sensed and recorded by the DCS. The four ducts combine to a single duct and the mixed gas temperature at this point is also sensed and recorded by the DCS.

WSA Condenser (E-209) Temperature switches are mounted in each of the inlet ducts of the condenser. These switches are used to detect sudden temperature decreases associated with sulfuric acid overflow into the inlet ducts which would indicate an acid system plug. The inlet and outlet flue gas pressures are measured and recorded by the DCS. The flue gas temperature exiting the WSA condenser is measured, and is controlled by the vane actuation of the cooling air fan.

Cooling Air Fan (K-230) The cooling air flow is monitored by the cooling air fan outlet thermal dispersion flow element and recorded by the DCS. The flow is varied by modulation of the cooling air fan vane actuator. The cooling air fan vanes are controlled by a Limitorque actuator. Vane positioning is based on WSA condenser flue gas outlet temperature and process inlet flow. Vane position is fed back to the DCS through a position transmitter. The fan discharge air temperature is monitored and is controlled through modulation of the cooling air recirculation damper actuator. The WSA condenser cooling air discharge is also measured. The fan outlet pressure is monitored and transmitted to the DCS. The motor for the cooling air fan is protected against over-temperature and excessive stop/start cycles by a motor protection relay in the motor control center. The motor run current is also monitored and indicated by the DCS.

Acid Holding Tank (B-220) The acid holding tank level (0-100%) is measured and transmitted to the DCS. The tank is also equipped with a redundant level switch that relays a signal to the DCS upon a 70% full level. The tank level is controlled by regulation of the product acid flow control valve which meters acid from the circulation loop to a storage tank. The recirculated acid flow rate is sensed and transmitted to the DCS.

Acid Circulation Pumps (P-221 A/B) The dilution water flow control valve receives a signal, representing a 0-5 gpm flow, from the DCS. The DCS controls the dilution water flow by sensing acid concentration. The acid circulation pump inlet temperature is sensed and recorded by the DCS. A temperature switch is provided to protect the pumps from high temperature operation. The acid circulation pumps are equipped with inlet and outlet isolation valves, outlet check valves, and drain valves. The pumps can only be started by the DCS through the MCC. The standby pump will automatically start on an in-service pump trip.

Acid Cooler (E-222) The acid cooler is equipped with a pressure gauge, gauge isolation valve, and drain valve at the acid inlet. The acid outlet temperature is sensed and recorded by the DCS.

The water inlet is equipped with an isolation valve, temperature gauge, pressure gauge, instrument isolation valve, and a drain valve. The water outlet is equipped with a temperature gauge. The water flow through the cooler is controlled by the DCS by monitoring the acid cooler acid outlet temperature and modulating the water flow control valve.

Acid Storage Tank (B-223A&B) The acid flow to the storage tanks is controlled through the product acid flow control valve by the DCS. Both storage tanks have isolation and drain valves. Acid Storage Tank A is equipped with a level element, which senses and transmits the tank level to the DCS and a transmitter for local indication of level. The off-spec tank is also equipped with a backup high-level switch that relays a signal to the DCS. Tank A overflows to tank B which has an overflow drain to the containment area. Both tank levels are also locally indicated in the acid transfer shed.

Air Compressors Two 100% capacity air compressors are supplied with the plant, one operating and one spare. Each unit is controlled by an air pressure switch located on the receiver tank. Automatic backup is accomplished by offsetting the pressure switch setpoints for each unit and placing both units in automatic mode. A low pressure switch is also provided at the discharge of each unit to alarm compressor malfunction. These switches are wired to a common annunciator at the baghouse control panel.

Load Control The mass flow of flue gas, or load, being treated by the SNOX plant is measured by the venturi flow monitor (V-101). A control loop in the DCS compares this measurement to a set point value and adjusts the position of control vanes for the flue gas booster fan accordingly. The control vanes are actuated by a Limitorque actuator. There are two pressure trips associated with this subsystem. These trips will protect the ductwork by initiating a plant trip if excessive pressure or vacuum is detected. The motor for the flue gas booster fan is protected against over-temperature and excessive stop/start cycles by a motor protection relay in the motor control center. The motor run current is also monitored and indicated by the DCS.

Continuous Emission Monitoring System The CEMS is an extractive type system manufactured by HORIBA. It provides SO_2, NO_x and O_2 signals at the process inlet; NO_x and O_2 signals after the SCR; and SO_2, NO_x, and O_2 signals at the WSA condenser outlet. These signals are used to monitor system and reactor removals of SO_2 and NO_x. The NO_x signals are also used to modulate NH_3 addition and thus control NO_x removal. All CEMS signals are continuously recorded by the DCS. A "CEMS Trouble" alarm will be displayed at the DCS in the event of any CEMS local trouble alarm.

4.2.2 Test Methods

As described in the October, 1991 Environmental Monitoring Plan (EMP) for the SNOX project, no compliance monitoring was required or conducted for this project. Both continuous and manual supplemental monitoring were conducted however, the results of which were published in the Quarterly Environmental Monitoring and Technical Reports. The results of the test program described in the Quarterly Reports are also included elsewhere in this report.

Continuous Supplemental Monitoring The DCS at the SNOX Demonstration Plant is designed to scan and record all instrument readings periodically during plant operation. The DCS recorded data was presented quarterly in the Environmental Monitoring Reports and discussed in the Quarterly Technical Progress Reports. The data captured and recorded on a routine basis was:

- Inlet pressure in in. w.g.
- Inlet temperature in deg. F.
- Unit 2 load percent
- SNOX load in lbs./min.
- Inlet NO_x in ppmv
- Inlet SO_2 in ppmv
- Inlet O_2 in vol. %
- Outlet NO_x in ppmv
- Outlet SO_2 in ppm
- Outlet O_2 in vol. %
- Discharge air temp. in deg. F.
- Acid mist in ppmv
- Product acid flow in gpm.

Manual Supplemental Monitoring To satisfy these requirements, a Data Collection Plan (DCP) was developed. The DCP is divided into individual test batteries that provide an overall description of the test, estimated duration, number of personnel required and the test method to be used. Additionally, where required a test matrix was provided. All data collected from the test batteries was compiled into a master data base along with any supplemental information.

Each test battery was assigned a priority rating based on its direct pertinence to the Demonstration objectives. Ratings of 1, 2, and 3 are deemed necessary for the fulfillment of these objectives. A priority rating of 4 was assigned that indicates that the data may be of value but is not deemed necessary for the fulfillment of the objectives.

In many instances, the explicit test method to be used is discussed in the testing description of the specific test battery. Where not explicitly stated all test procedures conformed to industry standards such as those by EPA, ASTM, EPRI, etc..

Table 4-4 lists stream sampled, frequency, schedule and method used to conduct the monitoring activities. A complete listing of the tests conducted in the DCP is provided in Appendix A.

Table 4-4 Monitoring Activity

Monitoring Point	**Parameters**	**Techniques**	**Frequency**	**Duration**
ESP Inlet	Flow	EPA 1 & 2	Once during Initial Plant Testing	Initial Testing only
	Particulate	EPA 5 or 17		
	SO_2, SO_3	MACS(1)		
	NO_x	EPA 7E		
	O_2, CO_2	EPA 3		
	NH_3	ISE (3)		
	Vanadium	EPA Multi-metals		

Stack	Flow	EPA 1 & 2	Once during Initial Plant Testing	Initial Testing only
	Particulate	EPA 5 or 17		
	SO_2, SO_3	MACS [1]		
	NO_x	EPA 7E		
	O_2, CO_2	EPA 3		
	NH_3	ISE [3]		
	Vanadium	EPA Multi-metals		
SNOX Inlet	Flow	EPA 1 & 2	Three times during Demonstration	Entire Demonstration Period
	Particulate	EPA 5 or 17		
	SO_2, SO_3	MACS[1]		
	NO_x	EPA 7E		
	O_2, CO_2	EPA 3		
	NH_3	ISE [3]		
	Vanadium	EPA Multi-metals		
	SO_2, NO_x, O_2	CEMS	Continuous	Life of Project
	Flow	Venturi		
	Temperature	Instrument		
	Pressure	Instrument		

Monitoring Point	**Parameters**	**Techniques**	**Frequency**	**Duration**
SNOX Outlet	Flow	EPA 1 & 2	Three times	Entire

	Particulate	EPA 5 or 17	during	Demonstration
	SO_2, SO_3	MACS [1]	Demonstration	Period
	NOx	EPA 7E		
	O_2, CO_2	EPA 3		
	NH_3	ISE [3]		
	Vanadium	EPA Multi-metals		
Condenser Out	SO_2, NO_x, O_2	CEMS	Continuous	Life of Project
	Acid Mist	Instrument		
Cooling Air	Temperature	Instrument	Continuous	Life of Project
Sulfuric Acid	Concentration	Titration	Monthly	Life of Project
	Concentration	Instrument	Continuous	
	Vanadium	AA [2]	Monthly	
	Total Flow	Instrument	Continuous	
	NO_3, NH_4	Colorimetric	Monthly	
	SO_2	Titration	Monthly	
Fly Ash	Vanadium	AA [2]	Quarterly	Life of Project
Siftings	Vanadium	AA [2]	Each Occurence	Each Occurence
	Mass	Spring Scale		
Coal Feed	H_2O, Ash, S, Vol., C, Btu/lb.	Proximate	Monthly	Life of Project
	C, H, N, O	Ultimate		

SO_2 Catalyst	Mass	Spring Scale	Each Occurence	Each Occurence

(1) Miniature acid - concentration system
(2) Atomic absorption
(3) Ion specific electrode

4.3 Analyses of Feed Stocks, Products and Reagents

As the SNOX process uses no reagents for SO2 removal, the analyses listed in this section are comprised of coal analyses and sulfuric acid product analyses. Analyses of the coal is listed in Tables 4-5 through 4-8, and the sulfuric acid concentrations are listed in Figures 4-1 through 4-6. A complete analysis of the acid product is given in Section 5.6, Table 5-23.

Table 4-5 Coal Analysis for 1991

Year	Month	HHV btu/lb	Moisture %	Ash %	Sulfur %
1991	Mar	11,718	7.93	11.81	2.91
	Apr	11,710	7.85	11.71	2.81
	May	12,003	6.81	10.76	2.76
	Jun	11,897	6.34	11.96	2.93
	Jul	11,768	6.99	12.14	2.88
	Aug	11,866	6.54	11.86	2.95
	Sep	12,024	5.43	12.31	2.72
	Oct	11,867	5.89	12.48	2.75
	Nov	11,917	6.63	11.51	2.81
	Dec	11,846	7.55	11.33	2.63
	Average	11,862	6.80	11.79	2.82

Table 4-6 Coal Analyses for 1992

Year	Month	HHV btu/lb	Moisture %	Ash %	Sulfur %
1992	Jan	11,829	7.42	11.80	2.80
	Feb	11,878	7.45	11.30	2.94
	Mar	11,998	7.81	10.41	2.65
	Apr	12,064	7.65	10.11	2.59
	May	12,066	7.29	10.34	2.74
	Jun	12,143	6.19	10.91	2.96
	Jul	12,001	7.17	10.63	2.85

	Aug	12,101	6.77	10.91	2.75
	Sep	12,074	7.03	10.97	2.72
	Oct	12,201	6.38	10.95	2.66
	Nov	11,835	7.04	11.92	3.07
	Dec	12,020	7.47	10.49	2.79
	Average	12,018	7.14	10.90	2.79

Table 4-7 Coal Analyses for 1993

Year	Month	HHV btu/lb	Moisture %	Ash %	Sulfur %
1993	Jan	11,892	7.97	11.18	2.77
	Feb	11,935	7.50	11.19	2.82
	Mar	11,685	9.22	10.91	2.70
	Apr	11,896	7.52	11.34	2.62
	May	11,920	6.69	11.58	2.63
	Jun	11,990	6.59	11.72	2.67
	Jul	12,175	5.89	11.16	2.60
	Aug	11,990	6.59	11.50	2.57
	Sep	12,108	6.60	11.27	2.69
	Oct	11,937	7.36	11.12	2.52
	Nov	11,734	8.32	11.55	2.74
	Dec	11,935	8.02	10.73	2.60
	Average	11,933	7.36	11.27	2.66

Table 4-8 Coal Analyses for 1994

Year	Month	HHV btu/lb	Moisture %	Ash %	Sulfur %
1994	Jan	11,718	9.36	10.78	2.64
	Feb	11,772	8.43	11.01	3.00
	Mar	11,731	8.32	11.72	2.99
	Apr	11,849	7.68	11.41	3.09
	May	11,983	6.63	11.75	3.00
	Jun	11,889	6.78	11.88	2.92
	Jul	11,954	6.74	12.07	2.68
	Aug	11,944	7.10	11.53	2.74
	Sep	11,920	6.77	11.48	2.76
	Oct	12,027	6.43	11.27	2.87
	Nov	12,165	6.59	11.08	2.92
	Dec	11,981	7.46	11.26	2.92

	Average	11,911	7.36	11.44	2.88

Figure 4-1 Acid Concentration, March thru June, 1992

Figure 4-2 Acid Concentration, Julythru September, 1992

Figure 4-3 Acid Concentration, October thru December, 1992

Figure 4-4 Acid Concentration, January thru March, 1993

Figure 4-5 Acid Concentration, April thru June, 1993

Figure 4-6 Acid Concentration, Octoberthru December, 1993

4.4 Data Analysis Methodology

Three forms of data were accumulated and analyzed during the course of the project: process operating, sampling, and laboratory analysis data. Responsibility for final data reduction and integration into final format rested with the project engineer.

Process operating parameter data (e.g., coal flows, load, air flows, steam temperatures) were obtained from ABB and plant personnel. The project engineer was responsible for assembling this data for use in data reduction and identification.

Sampling data recorded was first reduced by the sampling crew member performing the measurement. Field data was first entered directly into spreadsheet programs to determine velocities, sampled volumes at standard conditions, isokineticity, etc. All reduced field measurement data was transferred by the individual sampling team member to the test engineer. Continuous monitor data was recorded on a log form. A hard copy was obtained from the strip charts. The test engineer reviewed this data for completeness and performed selected calculation audits to ensure validity. The data was then assembled into a report that was submitted to the project engineer. This report contained:

- Sampling train spreadsheet printouts
- Continuous monitor data sheets, or analogous data acquisition printout and strip-chart recordings.

Laboratory analysis data were obtained from the organic, inorganic (trace metal), and feed characteristics (ultimate analysis) procedures performed. The analyst performed initial data reduction activities and prepared a report listing analyzed concentrations for each analyte in each sample for the laboratory supervisor. The laboratory supervisor reviewed this report and the raw analysis data for appropriate interpretation.

Chemical analysis data for the analyses performed in commercial laboratories (i.e. the trace metal analyses) were gathered by the laboratory supervisor and distributed to the project engineer and the QA Manager. This data was generally in the form of quantity of analyte measured per weight or volume of sample submitted (mg/kg or mg/L). The project engineer combined these data with the field-sampled volumes or weights, and reduced the raw analysis data to concentrations in the sample stream (e.g., :g/dscm).

The laboratory supervisor reviewed all reduced data from all analysts and also reviewed the field laboratory analysis report. This report states the analytical methods used and presents analytical results. Measured quantities in the aliquot analyzed (e.g., mg/L analyzed) were reported. Raw analysis data, in the form of notebook page copies, spectra, chromatograms, and strip charts, are kept indefinitely in the document archives.

The Test Engineer integrated all test-related reports (operation, sampling, and analysis reports), performed final data reduction, interpreted and analyzed the results, then drafted the test report, presenting measurement results, interpretations, engineering analysis results and conclusions. This report was reviewed for completeness, accuracy, and soundness of conclusions by the Test Contractor Project Manager before being submitted to ABBES.

Data validation and data reduction were performed at several levels in the project. At the most fundamental level, the test engineer and the laboratory supervisor reviewed and audited the measurement reports from individual samplers and analysts.

The laboratory supervisor audited all laboratory analysis reports, including those from other laboratories for completeness. Data validation was checked by comparing reported results to qualitative expectations. In addition, data validity was checked by reviewing reported results obtained on blanks, matrix spikes, and matrix spike replicate samples submitted for analysis.

The QA Manager was also responsible for performing selected data audits. In these, the entire analysis file for randomly selected files were reviewed, with checks for completeness of documentation, use of correct procedures, and appropriateness of data interpretation and reduction. All sampling and laboratory data was also reviewed. For example, log books, data logger outputs, continuous monitor outputs, sampling log books, analysis results, QC checks, and sample calculations were all areas of review.

Finally the test contractor project manager, in the review of the test report, performs the final level of data validation. The project manager compares reported findings with expectations or with findings reported in other efforts and verifies selected calculations using independent algorithms.

To ensure that the data collected was of known quality and accuracy, instrument and sampling equipment calibrations were accomplished in accordance with a set schedule.

The O_2, CO, CO_2, NO_x, and SO_2 CEMs were calibrated to standard reference procedures. Table 4-5 below summarizes the CEM calibration procedures and frequencies. Basically, however, each instrument was zeroed and spanned using certified calibration gases prior to the start of sampling. Calibration gases were introduced into the system ahead of the conditioning system. As part of the calibration and zeroing procedure, the linearity and drift of each instrument of each instrument is recorded. The values are then used to correct the data accordingly. If a calibration check indicated a greater than 5% change, the monitor was adjusted, and this adjustment noted in the test data documentation. The calibration gases used were certified for use with the respective reference methods cited in Table 4-9.

In the case of sampling trains, several variations of Method 5 or Method 17 were used in obtaining critical measurements for all tests. All probe nozzles, orifice meters, and dry gas meters utilized in variations of the Method 5 sampling trains, used to collect flue gas samples, particulate and HCl samples, were calibrated prior to the initiation of the test program. Probe nozzles were calibrated using the procedures and acceptance criteria outlined for Method 5 (40 CFR Part 60, Appendix A). The orifice and dry gas meters were calibrated relative to a bell prover (spirometer) that is NBS-traceable. This primary standard means of calibration provided an on-site reference dry gas meter to calibrate other instruments.

Table 4-9 Field Measurement Calibration

Measurement Parameter	Measurement Method	Calibration Method	Frequency of Calibration	Reference Standard
Continuous Emission Monitors				
O_2, CO_2	Paramagnetic	Zero & Span Midpoint calck	Before and After each test	Certified Calibration Gas
CO Monitor	Nondispersive Infrared (NDIR)	Zero & Span Midpoint calck	Before and After each test	Certified Calibration Gas
NO_x monitor	Chemilumines-cent	Zero & Span Midpoint calck	Before and After each test	Certified Calibration Gas
SO_2 monitor	UV	Zero & Span Midpoint calck	Before and After each test	Certified Calibration Gas
Sampling Train				
Velocity	S-type pitot tube	[illegible]	[illegible]	Standard S-type pitot tube
Pressure	Magnehelic gauge	Two-pt comparisonvs reference pt.	Every 6 months	Inclined Manometer
Temperature	Type K thermocouple	Two-pt comparisonvs reference pt.	Every 6 months	Mercury in glass thermometer
Gas sampling flow rate	Orifice meter	Comparison to reference std.	Before and after test program	Standard dry gas meter
Gas sampled volume	Dry gas meter	Comparison to reference std.	Before and after test program	Standard dry gas meter

4.5 Data Summary

This section presents operating data from the entire demonstration phase of the project. The information referenced herein is a compilation of data from the Quarterly Technical Reports and the Environmental Monitoring reports. It is presented in Appendix B. Tabular data from the above mentioned reports is listed in Tables B-1 through B-14; plant load information in Figures B-1 through B-8; and operating time and acid production in Figures B-9 through B-18.

4.6 Operability and Reliability

The major process equipment of the SNOX plant performed essentially without problems over the duration of the testing phase of the program. Problems which did occur and which interfered with plant operation involved auxiliary equipment.

During start-up of the plant in the first quarter of 1992, malfunctioning of the support burners and the flue gas booster fan caused the majority of down time. Three natural gas burners are used in the demonstration plant: Number 1 to heat the flue gas slip stream before entering the SNOX fabric filter; Number 2 is a support burner in the high temperature loop between the SCR and SO_2 reactor; and Number 3 is a reheat burner which heats the flue gas leaving the WSA condenser before it enters the flue gas stack.

Burners 1 and 3, from the same supplier, were very unstable in the flue gas atmosphere and tripped often, as well as having difficulty maintaining temperature. Several fixes were attempted by the supplier but these two burners were eventually replaced with burners of a different type, which have operated satisfactorily.

With respect to the flue gas booster fan, bearing overheating and failure occurred twice during early operation and required significant down time to repair. Although this fan does not see extremely high temperatures, it was decided to install a forced lubrication system for the bearings to produce cooler operating temperatures. This solution was implemented late in the testing program but has alleviated problems to date.

Some problems with the fabric filter also occurred during early operation. A "black light" test had been conducted before start up in March 1992 to identify any leaks in compartments or bags, and corrections had been made at that time to eliminate any leaks before the particulate tests were conducted. However, once into the test program, high inletloadings indicated that the condition of the fabric filter bags needed to be examined again. In subsequent filter bag examinations it was determined that the ash layer was significantly acidic, and that the bag material had lost much of its original strength. As a result, small pinholes were forming in some of the bags. A problem also existed with high pressure drop across the bags due to anuncleanable portion of deposit. It appears that this "sticky" layer of ash was the more acidic portion and was contributing to both the high pressure drop and bag deterioration.

As to the cause of the acidic ash, it appears that start up problems related to the natural gas fired, in-line burner upstream of the fabric filter contributed to periods of acid condensation occurring on the fly ash before the fabric filter or in the fabric filter. The purpose of this in-line burner is to both prevent the flue gas temperature from dropping below the aciddewpoint and to raise the flue gas to a temperature (. 400EF) which is typical of what would occur in a full size, integrated SNOX plant when the WSA condenser cooling air is used as combustion air to the boiler air preheater (raising the temperature of the flue gas exiting the airpreheater). Given the condition of the bags, it was decided to replace all of them during June of 1993.

With respect to the heat pipe gas/gas heat exchanger, minor temperaturemaldistribution was found to be present at the cold side outlet which feeds into the SCR. TheNO_x concentration distribution was determined to be very uniform as would be expected from the upstream gas path. Likewise, flow distribution at this location was consistent with what was expected from the gas path arrangement. The temperature profile, however, was not as predicted by the equipment manufacturer. This profile, which falls off significantly toward the wall opposite the partition plate, represented a potential catalyst fouling situation via ammonia compound condensation. To alleviate this potential, ammonia injection in the low temperature area was stopped. The SCR

NO_x removal efficiency was only minimally affected since this area was of low velocity and represented only a small fraction of the flue gas mass flow.

The SCR system itself is very passive and has provided reliable, maintenance free operation. One equipment selection hurdle, however, was encountered during start-up. The originally installed ammonia pump, which was of a diaphragm type, operated satisfactorily only at low ammonia flow rates. As the pump stroke was increased to provide additional flow, flashing occurred in the pump suction. Some modifications were made to the pump suction piping and ammonia storage tank operating level, but only minimal performance improvement was obtained. This problem was corrected by a change to a spur gear pump for ammonia pressurization. This pump selection has performed very well and allows the full range of design ammonia flow rates.

During normal operation of the plant, SCR system performance has been as expected with the exception of two developments resulting from the operation of other, upstream equipment. First, the low temperature zone at the inlet to the SCR reactor, discussed earlier, has necessitated the restriction of NH_3 addition in the affected area. This small, outboard zone, however, is a low mass flow region and has not affected NO_x removal significantly.

After about two years of operation a problem developed with the duct lining in the WSA condenser outlet duct to the Number 3 reheat burner. This duct was coated initially with an epoxy based coating but the coating deteriorated, exposing the A36 carbon steel ductwork. Since the outlet of the WSA condenser is at the sulfuric aciddewpoint, even the low mist concentration from the condenser (<5 ppm normally) results in the walls of the duct being wetted. In consulting the SNOX plant in Denmark it was confirmed that they had similar problems but were experiencing success with a vinyl ester coating. Although the Niles SNOX plant ductwork had little metal loss (measured after the duct was partially cleaned), it was decided to fabricate new ductwork and coat it with the vinyl ester based material before installation to eliminate considerable down time which would occur with further cleaning andrecoating of the existing ductwork in place.

4.6.1 Critical Component Failure Analysis

Critical components in the SNOX process are the SCR reactor, SO_2 converter, and WSA Condenser. In general, operating experience with these components was primarily trouble free. A problem related to the SO_2 Converter, although not with its process section, was failure of the expansion joints at the outlets of the converter. For the WSA condenser, performance was very good, except for a minor deterioration and leak at the flue gas inlet toward the end of the test program.

The converter expansion joint problems stem from the unavailability of materials able to withstand the 800EF gas temperatures, high SO_3 concentrations, required movement, and sealing qualities. Because conventional expansion joint material must be left uninsulated, the adjoining steel temperature can fall below the acid dewpoint and suffer acid attack. The initial joints employed a PTFE coated fiberglass material and were insulated on the outside to prevent condensation of sulfuric acid on the inner surface of the joints. However, temperatures were too high for the joint material and eventually caused failure. The second design employed was a metal foil joint which was also insulated on the outside. These joints could not handle the degree of mating flange movement and the foil ripped soon after installation.

This overall problem with these expansion joints also occurred at the NEFO SNOX plant in Denmark and was ultimately solved by use of an air purged joint. A small quantity of hot air from the WSA condenser is employed for this purpose at NEFO and the same design was incorporated into the Niles SNOX plant. Future designs will incorporate this type of expansion joint although a metal joint is also a possibility if the structures and ducts have been initially designed to produce movements compatible with metal expansion joints.

Toward the end of the testing period in 1994, an acid leak developed at one of the flue gas inlets to the WSA condenser. On inspection it was determined that flue gas was leaking past the joint between acid brick and the metal flange at the inlet nozzle. Condensation of acid in the nozzle beneath the acid brick perforated the nozzle and caused the leak. Nozzle reinforcement material was welded onto the existing nozzles and the joint between the brick and flange was sealed with PTFE rope andregrouted.

A redesign of this section has also been accomplished byHaldor Topsoe, and will be incorporated into future designs. The new design offers a better method of sealing between the acid brick and the metal inlet flange.

5.0 TECHNICAL PERFORMANCE

5.1 Unit 2 Characterization

This section covers the Test Group 2A environmental sampling and analytical results. This test group focused on characterizing the process conditions and emissions from the host boiler, Unit 2 of the Niles Station of Ohio Edison. Test Group 2A is part of the Battery 2 Baseline testing to quantify reference environmental conditions on which to compare the performance results of the SNOX process. Test Group 2A had the objective of characterizing the existing Ohio Edison Niles Plant Unit 2 ESP inlet and outlet process streams to determine the types and quantities of emissions in the feed stream to the SNOX Demonstration Plant.

Results from the Battery 2A test series conducted on the ESP inlet and outlet are reported in this section. Stack gas flow, temperature, moisture content, and static pressure were taken during most of the test runs mentioned above and are reported with the particulate test results.

Particulate emissions test runs (in triplicate) incorporating flue gas flow, temperature, moisture, O_2, and CO_2 were performed at the ESP inlet and outlet. Tables 5-1 and 5-2 summarize the test results.

Parameter	Run 1	Run 2	Run 3	Average
Flowrate,acfm	327,000	314,000	308,000	316,000
Flowrate,dscfm	212,000	201,000	202,000	205,000
Velocity, ft/sec	48.1	46.3	45.4	46.6
Temperature, EF	287	287	285	286
% H_2O	8.0	9.5	7.5	8.3
% O_2	5.5	5.5	5.5	5.5
% CO_2	14.0	14.0	14.0	14.0
Particulate concentrationgr/dscf	1.06	0.841	1.09	0.9975

Parameter	Run 1	Run 2	Run 3	Average
Flowrate,acfm	250,000	252,000	243,000	249,000
Flowrate,dscfm	171,000	168,000	163,000	167,000
Velocity, ft/sec	43.5	44.2	42.4	43.4
Temperature, EF	260	271	269	266
% H_2O	6.6	7.6	7	7.1
% O_2	5.5	5.5	5.5	5.5
% CO_2	14.0	14.0	14.0	14.0
Particulate concentrationgr/dscf	0.0333	0.0063	0.0058	0.0151

The flow rate results demonstrated that generally good uniformity existed between the triplicate runs at a station, but there was a considerable discrepancy of 23% between the upstream and downstream stations. It is suspected that some unquantified portion of the flue gas flow was going to the Demonstration Plant during the testing, therefore bypassing the ESP. The ESP collection efficiency during the particulate test series averaged 98.5%. This efficiency estimate is based on upstream and downstream particulate concentrations during the three sets of particulate runs rather than on mass emission rates which are possibly biased by the bypass flue gas.

Particle sizing runs were performed in triplicate at the ESP inlet and outlet utilizing California Air Resources Board (CARB) Method 501 (Andersen multi-stage inertial impactor). Table 5-3 presents the particle size fractions for the 9 size cuts, upstream and downstream of the ESP. The size distribution showed generally good agreement for the triplicate replication runs. As expected, the particle size distribution was strongly biased to the smaller size ranges by the effect of the ESP.

Table 5-3 Particle Sizing (CARB Method 501) Test Results (microns)

ESP Inlet	Run 1			Run 2			Run 3			Average		
Stage	%[a]	CPLT[b]	Diam.	%	CPLT	Diam.	%	CPLT	Diam.	%	CPLT	Diam.
Precutter	66.4	33.6	>10.0	74.3	25.7	>10.0	75.1	24.9	>10.0	71.9	28.1	>10.0
0	1.7	31.9	6.69	3.4	22.3	6.57	2.0	22.9	6.58	2.4	25.7	6.61
1	3.5	28.4	6.33	4.7	17.6	6.21	3.4	19.5	6.23	3.9	21.8	6.26
2	7.7	20.7	3.86	4.1	13.5	3.79	4.5	15.0	3.80	5.4	16.4	3.82
3	8.5	12.2	2.68	4.5	9.0	2.63	4.0	11.0	2.64	5.7	10.7	2.65
4	3.1	9.1	1.38	3.4	5.6	1.36	4.3	6.7	1.36	3.6	7.1	1.37
5	2.1	7.0	0.68	1.5	4.1	0.67	1.1	5.6	0.67	1.6	5.6	0.67
6	2.5	4.5	0.45	0.8	3.3	0.44	1.4	4.2	0.44	1.6	4.0	0.44
7	2.8	1.7	0.25	1.6	1.7	0.24	2.0	2.2	0.25	2.1	1.9	0.25
Backup	1.6	0.1	<0.25	1.6	0.1	<0.24	2.2	0.0	<0.25	1.8	0.1	<0.25

ESP Outlet	Run 1			Run 2			Run 3			Average		
Stage	%	CPLT	Diam.	%	CPLT	Diam.	%	CPLT	Diam.	%	CPLT	Diam.
Precutter	12.9	87.1	>10.0	31.7	68.3	>10.0	39.7	60.3	>10.0	28.1	71.9	>10.0
0	12.6	74.5	6.41	9.2	59.1	7.08	10.5	49.8	5.50	10.8	61.1	6.33
1	11.4	63.1	6.06	9.6	49.5	6.69	7.7	42.1	5.20	9.6	51.6	5.98
2	6.2	56.9	3.70	6.1	43.4	4.09	4.2	37.9	3.17	5.5	46.1	3.65
3	6.1	50.8	2.57	3.5	39.9	2.83	3.3	34.6	2.20	4.3	41.8	2.53
4	0.7	50.1	1.32	4.6	35.3	1.46	5.0	29.6	1.13	3.4	38.3	1.30
5	2.2	47.9	0.65	4.9	30.4	0.72	4.7	24.9	0.55	3.9	34.4	0.64
6	7.8	40.1	0.43	10.7	19.7	0.47	9.9	15.0	0.36	9.5	24.9	0.42
7	16.8	23.3	0.24	15.1	4.6	0.27	10.7	4.3	0.20	14.2	10.7	0.24
Backup	23.4	0.0	<0.24	4.6	0.0	<0.27	4.3	0.0	<0.20	10.8	0.0	<0.24

[a] Percent of Total Particulate
[b] Cumulative Percent Less Than Stated Size

5.2 Particulate Removal

Results from the Battery 2B calibration test series are reported in this section along with additional particulate sampling in Batteries 3 and 4.

Particulate emissions tests, initially at 100% load condition, were performed in triplicate at the baghouse inlet and outlet. Table 5-4 summarizes the test results. The baghouse was operating at an average efficiency of 98.6%.

EPA Method 17 was used for baghouse inlet particulate determinations. The flue gas was sampled isokinetically with an in-stack filter. Table 5-5 presents data from the three baghouse inlet runs. The average inlet particulate concentration was 0.65 grains per dry standard (68EF, 29.92 in. Hg.) cubic foot of flue gas. The average particulate emission rate was 440 lb/hr. The particulate loading results exhibited considerable variability, particularly for run 2. The variability is outside the range typically experienced from routine variations in coal quality and boiler load, but are within the range which could result from a higher incidence of soot blowing on Unit 2.

EPA Method 5 was used for the determination of particulate loading at the baghouse outlet. Table 5-6 presents the run data. Particulate data reported includes front half catch only (i.e., no condensibles). Baghouse outlet runs were conducted for approximately 8 hours in order to obtain a sufficiently large sample for quantification given the low downstream particulate concentrations. The average particulate concentration at the baghouse outlet was 0.0089 gr/dscf of flue gas. The average particulate emission rate was 5.97 lb/hr. The outlet particulate loadings showed variability, but these variations generally correlated with the changes in inlet particulate loading, particularly for run 2.

Additional particulate emissions tests were performed later in the program at the 75%, 100%,

Run No.	**Emission Rate lb/hr**		**Concentration gr/dscf**		**Removal Eff %**
	Inlet	**Outlet**	**Inlet**	**Outlet**	
1	306	3.11	0.46	0.0047	98.98
2	600	9.02	0.88	0.0133	98.50
3	413	5.77	0.61	0.0087	98.60
Avg	440	5.97	0.65	0.0089	98.64

and 110% load conditions at the baghouse inlet (system inlet), the baghouse outlet, and the system outlet. Test runs were 80 minutes in length at the baghouse inlet (two runs were performed per one baghouse outlet/system outlet run) and 8 hours in length at the baghouse outlet and system outlet. Test results are summarized in Table 5-7.

Parameter	Run 1	Run 2	Run 3	Average
Flue Gas				
Volume sampled, dscf	38.21	39.20	36.29	**37.90**
Average velocity, fps	71.3	73.5	73.0	**72.6**
Average temperature, EF	413	418	408	**413**
Average flowrate, dscfm	77,334	79,705	78,947	**78,662**
% CO, dry	14.0	14.0	14.0	**14.0**
% O_2, dry	4.0	4.0	4.0	**4.0**
% H_2O	10.3	8.5	9.2	**9.3**
Particulate				
Concentration, gr/dscf	0.46	0.88	0.61	**0.65**
Emission rate, lb/hr	306	600	413	**440**

Parameter	Run 1	Run 2	Run 3	Average
Flue Gas				
Volume sampled, dscf	309.78	255.8	251.8	272.5
Average velocity, fps	69.3	71.4	70.4	70.2
Average temperature, EF	383	394	393	390
Average flowrate, dscfm	77,667	79,225	77,452	78,115
% CO, dry	13.0	13.0	13.0	13.0
% O_2, dry	4.0	4.0	4.0	4.0
% H_2O	8.9	9.8	9.6	9.4
Particulate				
Concentration, gr/dscf	0.0047	0.0133	0.0087	0.0089
Emission rate, lb/hr	3.13	9.02	5.77	5.97

Baghouse collection efficiency averaged 98.55% for the 75% load runs, 98.66% for the 100% load runs, and 98.29 for the 110% load runs as seen in Table 5-8. The flyash particulate concentration downstream of the system outlet would be expected to be low due to the scrubbing action of the WSA condenser which captures small particulate in the process of condensing SO_3 and H_2O (small particles are used as condensation nuclei and end up in the product acid). The material caught on the particulate run filters is likely to be sulfuric acid mist and corrosion products from the sampling equipment. The Method 5 filters at this location were noted by the operators to be discolored consistent with acid/condensation nuclei carryover. The SO_3 concentration at the WSA outlet (determined by MACS sample runs from another test battery) was about 0.0004 gr/dscf. These levels, combined with condensation nuclei, would give residue levels which are a significant fraction of the measured particulate concentrations at that location.

Plant Load (%)	**Baghouse Inlet**[a]		**Baghouse Outlet**[a]		**System Outlet**[a]	
	gr/dscf	**lb/hr**	**gr/dscf**	**lb/hr**	**gr/dscf**	**lb/hr**
75	0.529	242	0.0082	3.50	0.0049	2.88
100	0.770	450	0.0107	6.02	0.0060	4.31
110	0.530	324	0.0089	5.54	0.0026	2.05

[a] Baghouse inlet results are the average of six runs and baghouse outlet and system outlet results are the average of three runs.

Run No.	**Particulate loading (lb/hr)**			**Removal efficiency (%)**	
	Baghouse inlet	Baghouse outlet	System outlet	Baghouse	System
3A1	242	2.61	2.01	98.9	99.1
3A1B	218	C	C	C	C
3A2	232	4.67	2.70	97.6	98.6
3A2B	149	C	C	C	C
3A3	159	3.24	3.95	98.9	98.7
3A3B	453	C	C	C	C
Average	**242**	**3.50**	**2.88**	**98.5**	**98.8**
3B1	419	3.24	2.40	99.3	99.5
3B1B	465	C	C	C	C
3B2	408	1.92	8.32	99.6	98.3
3B2B	572	C	C	C	C
3B3	410	12.9	2.22	96.9	99.5
3B3B	427	C	C	C	C
Average	**450**	**6.02**	**4.31**	**98.6**	**99.1**
3C1	315	6.43	1.41	98.0	99.6
3C1B	321	C	C	C	C
3C2	372	7.27	2.23	97.9	99.3
3C2B	302	C	C	C	C
3C3	392	2.91	2.52	99.1	99.2
3C3B	238	C	C	C	C
Average	**324**	**5.54**	**2.05**	**98.3**	**99.4**

In summary, performance of the PTFE fabric filter was not as good as expected (.0004 - .0008 gr/dscf with PTFE bags), although accurate sampling at this very low level of particulate is difficult. The presence of acid condensation or large particles may have skewed some of the test results. Rate of increase in pressure drop of the SO_2 converter, which collects additional particulate, was in line with lower loadings than were typically measured leaving the fabric filter. Otherwise, more frequent cleaning of the catalyst would have been necessary.

5.3 NO_x Reduction

Nitrogen oxides are converted to nitrogen and water vapor in the SNOX process via selective catalytic reduction with ammonia. The catalyst and SCR reactor design used for this project were supplied by Haldor Topse A/S. This design is a top down gas flow arrangement with three catalyst bed levels, two of which are initially filled and one is spare. The reactor casing is constructed of A-204 high temperature steel and sized for an effective design space velocity of about 7500 h^{-1}.

The strategic location of the SCR reactor in the SNOX process as compared to conventional high dust SCR applications results in several benefits. First, the post fabric filter, low dust environment allows the use of high specific area catalyst and thus lower catalyst volumes; sootblowers are not necessary; and flyash contamination with ammonia is avoided. In addition, much lower catalyst erosion can be expected as well as less potential for poisoning from gaseous arsenic, significantly increasing catalyst lifetime.

Second, the location of the SCR reactor upstream of the SO_2 converter allows operation at an NH_3/NO_x ratio (stoichiometry) in excess of 1.0 without the potential of downstream ammonium sulfate and bisulfate condensation which can be a result of excess ammonia slip. All ammonia slip in the SNOX process is oxidized in the downstream oxidation reactor. Operation with this ammonia surplus greatly reduces the catalyst volume necessary for a given NO_x removal.

The relative location of the two reactors has one other benefit. In conventional SCR applications, catalysts are required not to oxidize more than about 1% of the inlet SO_2 to SO_3 in order not to increase the downstream sulfuric acid dewpoint significantly. This requirement often has a side effect of reduced catalyst NO_x removal activity and thus higher catalyst volumes. The SNOX Process does not have this limitation since any SO_2 oxidation in the SCR reactor only benefits the oxidation reactor downstream.

The DNX-932 catalyst used in this design is a high activity, titanium oxide based monolithic type which operates in the temperature range of 650-800EF. This low particulate loading version of the DNX line has a hydraulic diameter of 0.122 in. and a specific area of 235 ft^2/ft^3.

All of the tests discussed in this section were performed at 100% unit load with an SCR inlet temperature of 715EF. NO_3/NH_x ratios were varied between 0.90 and 1.10 during the course of each test battery. A full SCR traverse for ammonia and sample point velocity was performed at the nominal 1.02 NH_3/NO_x ratio for each subgroup. Inlet and midpoint MACS sampling runs for SO_2 and SO_3 were performed for each subgroup only at the nominal 0.90 NH_3/NO_x ratio.

Test results from the Battery 5A parametric testing are presented in Table 5-9. The NO_x concentration at the SCR combined outlet (system midpoint) was lower than expected for all values of NH_3/NO_x ratio. SCR NO_x removal efficiencies for the individual NH_3/NO_x ratios are all greater than 99%.

The Group 5B parametric results are summarized in Table 5-10. The results are different from Group 5A; the NO_x concentration at the SCR outlet tends to decrease (as expected) as the NH_3/NO_x ratio increases. SCR NO_x removal efficiencies for the individual NH_3/NO_x ratios are greater than 99% except for the NH_3/NO_x = 0.90 result.

Battery 5C parametric results are summarized in Table 5-11. As in the other Battery 5 tests, the results generally conform to expectations; the NO_x concentration at the SCR outlet tends to decrease with increasing NH_3/NO_x ratio. SCR No_x removal efficiencies for the individual NH_3/NO_x ratios were generally greater than 95% (across the SCR reactor).

Run No.	SCR Outlet					System Inlet	
	Actual NH_3/NO_x	**NH_3 (ppmv)**	**Sample Vol (L)**	**NO_x (ppmv)**	**O_2 (%)**	**NO_x (ppmv)**	**O_2 (%)**
5A2-1	0.95	0	984.2	2	5.4	569	N/O
5A2-2	0.95	0	998.9	2	5.4	569	N/O
5A3-1	0.98	0	484.9	1	5.4	814	5.4
5A3-2	0.98	0	499.0	1	5.4	814	5.4
5A4-1	1.00	34.8	478.6	1	5.1	819	5.0
5A4-2	1.00	1.1	486.8	1	5.1	819	5.0
5A6-1	1.05	68.3	512.0	2	5.3	844	5.3
5A6-2	1.05	66.2	505.2	2	5.3	797	5.2
5A7-1	1.10	109.9	511.8	0	5.1	640	5.5
5A7-2	1.10	44.0	507.5	0	5.1	660	5.4
5A1-2	0.90	16.5	489.1	1	5.3	690	5.2
5A1-3	0.90	11.2	497.3	1	5.3	690	5.2

N/O = Monitor inoperative during this time period

Run No.	SCR Outlet					System Inlet	
	Nominal NH_3/NO_x	**NH_3 (ppmv)**	**Sample Vol (L)**	**NO_x (ppmv)**	**O_2 (%)**	**NO_x (ppmv)**	**O_2 (%)**
5C1-1	0.90	5.0	474.9	30[a]	6.8[a]	550	5.7
5C1-2	0.90	2.0	479.2	30[a]	6.8[a]	550	5.7
5C2-1	0.95	4.0	977.3	50[a]	6.4[a]	641	6.4
5C2-2	0.95	2.7	973.1	46[a]	6.2[a]	641	6.2
5C3-1	0.98	30.5	960.9	34[a]	6.1[a]	623	6.0
5C3-2	0.98	44.0	965.2	31[a]	6.1[a]	611	5.9
5C4-1	1.00	96.2	965.2	29[a]	6.3[a]	562	6.1
5C4-2	1.00	64.7	949.9	31[a]	6.3[a]	508	5.7
5C5-1	1.02	30.4	1036.0	1	5.8	628	5.4
5C5-2	1.02	9.1	1056.4	3	6.2	682	5.5
5C5-3	1.02	6.3	1043.3	1	N/O	670	5.6
5C5-4	1.02	12.1	1065.7	0	N/O	652	5.3
5C5-5	1.02	9.4	1071.4	14	N/O	748	5.3
5C6-1	1.05	32.1	952.4	33[a]	6.0[a]	529	5.4
5C6-2	1.05	30.9	948.2	32[a]	6.2[a]	533	5.5
5C6-3	1.05	41.9	1056.9	0	N/O	628	5.1
5C6-4	1.05	29.6	988.9	1	6.4	623	4.9
5C7-1	1.10	41.2	1060.9	0	N/O	642	5.4
5C7-2	1.10	45.6	1041.0	0	N/O	654	4.9

[a] System outlet monitored at this time
N/O = O_2 monitor not operating at this time

Run No.	SCR Outlet					System Inlet	
	Nominal NH_3/NO_x	**NH_3 (ppmv)**	**Sample Vol (L)**	**NO_x (ppmv)**	**O_2 (%)**	**NO_x (ppmv)**	**O_2 (%)**
5B1-1	0.90	1.7	1034.0	15.2	N/O	N/O	N/O
5B1-2	0.90	6.0	1050.7	15.2	N/O	N/O	N/O
5B2-1	0.95	30.3	1021.2	0.7	5.9	515	4.9
5B2-2	0.95	25.8	1049.8	0.7	5.9	515	4.9
5B3-1	0.98	35.3	1000.3	0.7	5.5	525	4.7
5B3-2	0.98	31.7	1066.2	0.7	5.5	525	4.7
5B4-1	1.00	26.7	1011.9	1.0	5.4	463	4.9
5B4-2	1.00	34.0	1039.1	1.0	5.4	463	4.9
5B6-1	1.05	104.5	526.2	1.0	5.3	511	4.4
5B6-2	1.05	11.7	512.6	1.0	5.3	511	4.4
5B7-1	1.10	44.7	507.2	1.3	5.1	592	4.8
5B7-2	1.10	37.2	533.8	1.3	5.1	592	4.8

N/O = Monitor inoperative during time period

In summary, data from Battery 5A appear to have unrealistically low NO_x concentrations exiting at the lower ammonia stoichiometries. The other two test series, 5B and 5C, are more representative of the relationship between ammonia stoichiometry and NO_x leaving the reactor. During Test Group 5C, system outlet concentrations of NO_x were also recorded for some of the tests and resulted in system removal efficiencies in the 92 to 94% range overall.

The SO_2 oxidation rate results (series 5S) for Test 1 are summarized in Table 5-12. MACS samples were taken at the system inlet and at the system midpoint. At a nominal full load flowrate of 69,000 dscfm of flue gas, the SO_2 oxidation rate is approximately 38 lb/hr as SO_2, or 2.7% oxidation based on inlet SO_2 concentration.

Table 5-13 presents the SO_2 oxidation rates (series 5S) for the 5B Group. MACS samples were taken at the system inlet and at the midpoint. At a nominal full load flowrate of 69000 dscfm of flue gas, the SO_2 oxidation rate is approximately 41 lb/hr as SO_2, or 2.9% oxidation.

Table 5-14 presents SO_2 oxidation data for Group 5C. MACS samples were taken at the system inlet and at the midpoint. At a nominal full load flowrate of 69,000 dscfm of flue gas, the SO_2 oxidation rate is approximately 38 lb/hr as SO_2, or 3.8% oxidation at the lower inlet SO_2 of 1440 ppm.

Ammonia emissions testing was performed at the 110% load condition at the baghouse outlet and system outlet on February 8, 1993. Identical tests were performed for the 100% and 75% load conditions on February 9, 1993. Bay Area Air Quality Management District (BAAQMD) Test

Run No.	**Sample Volume (dscf)**	**Wt-SO_2 (g)**	**SO_2 (ppmv)**	**Wt-SO_3 (g)**	**SO_3 (ppmv)**
Inlet 1	7.61	1.7187	3002[a]	0.0158	22.1[a]
Inlet 2	6.72	1.1269	2229	0.0058	9.2
Average	—	—	**2229**	—	**9.2**
Mid 1	6.66	1.0099	2015	0.0429	68.4
Mid 2	8.93	1.3533	2014	0.0499	59.4
Average	—	—	**2015**	—	**63.9**
Δ ppm SO_3					**54.7**
% SO_2 oxidation					**2.7**

[a]Suspected probe contamination; not used in calculation

Date	Run Load (%)	Run No.	Baghouse Outlet NH_3 (ppmv)	Baghouse 3-Run Average (ppmv)	System Outlet NH_3 (ppmv)	System Outlet 3-Run Average (ppmv)
2/8/93	110	1	(a)	C	0.4	C
2/8/93	110	2	0.0	0.0	0.8	0.8
2/8/93	110	3	0.0	C	1.3	C
2/9/93	75	1	6.4	C	0.0	C
2/9/93	75	2	7.0	6.9	0.0	0.0
2/9/93	75	3	7.2	C	0.0	C
2/9/93	100	1	8.6	C	(a)	C
2/9/93	100	2	7.9	8.5	0.0	0.0
2/9/93	100	3	9.0	C	0.0	C

(a) Unknown contamination caused cloudy, unusable sample

Run No.	Sample Volume (dscf)	Wt SO_2 (g)	SO_2 (ppmv)	Wt SO_3 (g)	SO_3 (ppmv)
Inlet 1	4.39	0.6704	2029	0.0844	10.6
Inlet 2	4.86	0.7490	2048	0.0017	3.7
Average	**C**	**C**	**2039**	**C**	**7.2**
Mid 1	4.28	0.9673	2972	0.0280	69.5
Mid 2	3.33	0.0832	332[a]	0.0184	62.0
Average	**C**	**C**	**2972**	**C**	**65.8**
) ppm SO_3					**58.6**
% SO_2 oxidation					**2.9**

[a] Suspect sample; not used in SO_2 average

Run No.	Sample Volume (dscf)	Wt SO_2 (g)	SO_2 (ppmv)	Wt SO_3 (g)	SO_3 (ppmv)
Inlet 1	3.56	0.3892	1453	0.0070	21.1
Inlet 2	3.54	0.3799	1426	0.0065	19.5
Average	**C**	**C**	**1440**	**C**	**20.3**
Mid 1	4.63	0.5390	1550	0.0355	81.5
Mid 2	4.65	0.5496	1571	0.0301	68.8
Average	**C**	**C**	**1560**	**C**	**75.1**
) ppm SO_3					**54.8**
% SO_2 oxidation					**3.8**

Method ST1-B was used. Results are presented in Table 5-15. There is a possibility that ammonia injection for fly ash conditioning upstream of the Ohio Edison Unit 2 ESP may have affected the ammonia concentration at the baghouse outlet (SNOX system inlet).

5.4 SO_2 Oxidation

Sulfur dioxide removal in the SNOX process is controlled by the efficiency of the SO_2 to SO_3 oxidation which occurs as the flue gas passes through the oxidation catalyst beds. The SO_2 converter, which contains the catalyst, is a vessel constructed of high temperature, carbon steel containing four panels installed in parallel, each with two vertical beds. The beds are filled with Haldor Topse VK-WSA sulfuric acid catalyst. Excess amounts of catalyst exist in the top and bottom of the converter. The flue gas is distributed uniformly over the eight catalyst beds through five inlet nozzles. After passing through the catalyst beds, the flue gas is discharged through four outlet nozzles. The catalyst is a vanadium-based oxidation catalyst in the shape of 0.4 inch O.D./0.16 inch I.D. x 0.35 inch rings.

Oxidation efficiency through the catalyst beds is controlled primarily by two factors - space velocity governs the amount of catalyst which is necessary at design flue gas flow conditions, and gas and bed temperature must be high enough to "ignite" or activate the SO_2 oxidation reaction.

Flue GasFlowrates in SO_2 Converter Outlet Ducts				
	Volumetric Flow		Velocity	Deviation
Duct	acfm	dscfm	fps	%
A	47,216	18,896	42.2	+ 0.73
B	48,608	19,458	43.4	+ 3.7
C	46,796	18,902	41.8	+ 0.76
D	43,944	17,777	39.2	- 5.2
Total	186,553	75,036		
Average	46,638	18,759	41.6	

In the tests conducted, temperature and flow measurements have been taken for the four catalyst panels. In order to have uniform space velocity for each panel, the flue gas flow to each must also be uniform. Because an in-line, trim burner is used at the Niles SNOX plant, temperature distribution entering the panels was measured since it can vary with this type of heat source. A tubular heater can also be employed in this location and will minimize the possibility of temperature maldistribution.

Table 5-16 lists results from the flow measurements and Table 5-17 contains the results from the temperature measurements. The flow measurements were made at the four outlet nozzles or ducts and temperature measurements were done at the five inlet nozzles or ducts. Flow through the four. catalyst panels is acceptably uniform and all quantities are within 6% of the average. With respect to temperature, it is also uniform and the temperatures in the five inlet ducts are within about 4% of the average.

Test Battery 6 was conducted to characterize and document the SO_2 oxidation efficiency of the SO_2 converter catalyst beds. In this test battery, the catalyst bed temperature was varied over the range of 715EF to 800EF at three different loads (80%, 100%, and 110% of design). Results from these tests are shown in Table 5-18. Gas flow was varied for the three loads from about 50,000 scfm to 90,000scfm.

Duct	Temperature Deg F	Deviation %
A	806	- 0.62
B	798	- 1.6
C	844	+ 4.1
D	801	- 1.2
E	804	- 0.86
Average	811	

For most of the data, the trend of higher SO_2 removal efficiency at higher catalyst bed temperature and lower gas flow (lower space velocity) is apparent

5.5 Sulfuric Acid Condensation

After the flue gas has passed through the SO_2 converter which has oxidized greater than 95% of the incoming SO_2 to SO_3, the gas must be cooled to induce the condensation of sulfuric acid. This cooling must be performed strategically as high SO_3 concentrations can represent a very aggressive atmosphere should condensation occur in the presence of unsuitable materials of

Test No.	Conv Temperature (EF)	WSA Flow (scfm)	System Outlet		Conv. In. SO_2 (ppmv)	SO_2 Rem Efficiency (%)
			CEMS SO_2 (ppmv)	MACS SO_3 (ppmv)		
6A9/7A12	834	88,900	60	6.2	1503	96.0
6A1/6T1/7A9	839	85,000	68	7.1	1584	95.7
6A1/6T1/7A9	838	83,900	67	6.4	1584	95.8
6A5/6T3/7A6	832	67,000	70	12.2	1652	95.8
7A3	816	50,600	47	14.4	1413	96.7
6A10/7A11	807	90,400	102	0.6	1413	92.8
6A2/7A8	807	86,600	87	1.0	1449	94.0
6A6/7A5	801	70,800	72	0.4	1449	95.0
7A2	789	52,700	66	1.2	1553	95.8
6A11/7A10	780	91,300	105	0.5	1655	93.7
6A3/6T2/7A7	791	84,200	100	0.6	1674	94.0
0A7/6T4/7A4	775	70,000	83	6.0	1500	95.5
7A1	762	60,900	73	24.1	1426	94.9
6A8	737	66,400	146	4.0	1426	89.7
7H3	811	87,700	91	13.2	1523	94.0
7H2	824	88,300	86	13.8	1483	94.2

construction. Depending on the actual concentrations of SO_3 and water, the acid dewpoint of this flue gas stream is in the range of 400EF.

Cooling the gas is performed through two pieces of equipment - first the hot side of the GGH and then the WSA condenser. As heat is transferred to the SCR reactor inlet stream via the GGH, the gas cools to about 510EF. This temperature change drives most of the SO_3 to hydrate with the available water to form H_2SO_4 vapor. The precooled gas enters the bottom of the condenser, then flows up through the interior of borosilicate glass tubes. Ambient air is passed across the exterior of the glass tubes counter currently to the flue gas flow. In this manner the flue gas is cooled to about 210EF and the cooling air is heated to 350E to 400EF. During the flue gas cooling, sulfuric acid vapor condenses in a filmwise fashion on the inner walls of the tubes and drains into the acid collection trough in the bottom of the condenser.

WSA SO_3 condensation efficiency can be calculated from a combination of source sampling results, coal sulfur analysis, acid production data taken from the SNOX DCS, and acid strength

Run No.	Calculated Inlet SO_3, (ppm)	Acid Production Rate (gal/hr)	SO_3 Condensation Rate (lb/hr)	Outlet SO_3, ppm (MACS)	SO_3 Condensation Efficiency (%)
7A12	1471	139.4	1625	6.2	99.6
7A9	1557	139.6	1646	7.1	99.5
7A9	1713	148.4	1786	6.4	99.6
7A6	2222	154.0	1848	12.2	99.5
7A3	1713	89.6	1072	14.4	99.2
7A11	1413	133.9	1595	0.6	100.0
7A8	1602	145.3	1731	1.0	99.9
7A5	1349	98.5	1192	0.4	100.0
7A2	2101	114.9	1382	1.2	99.9
7A10	1636	160.3	1864	0.5	100.0
7A7	1999	182.9	2100	0.6	100.0
7A4	N/A	N/A	N/A	6.0	N/A
7A1	N/A	N/A	N/A	24.1	N/A
7H3	1365	124.2	1480	13.2	99.0
7H2	1641	151.8	1794	13.8	99.2

N/A - No acid production data available from DCS laboratory results. WSA SO_3 condensation efficiency cannot be reported directly from source sampling results alone because of the lack of SO_2 and SO_3 concentrations at the condenser inlet. The condenser inlet location proved untestable due high temperature and SO_3 concentration which led to rapid probe failure. The computed efficiency was based on several assumptions. The first is that NH_3 carrier air and second support burner combustion air adds 20 % volume of flue gas dilution between the system inlet and the SO_2 converter inlet. This factor is necessary to correct system inlet CEMS SO_2 results to converter inlet conditions. The second is that there is no significant change in SO_2 concentration across the WSA condenser (no SO_2 oxidation). The third is that the acid production rate meter is accurate even though a field calibration was not performed. The fourth is that there is no leakage in the GGH. The final assumption is that the oxygen consumed in the SO_2 converter and the water consumed in the WSA condenser does not significantly affect flue gas composition.

Table 5-19 presents these results. Most of the data yields condensation efficiencies above 99.5%, with corresponding SO_3 outlets below 7ppm.

During particulate testing which was conducted in Test Batteries 3 and 4, manual samples were also taken for SNOX system inlet and outlet SO_2. MACS testing for SO_2 and SO_3 was performed at the baghouse inlet and system outlet at the 75% load condition, the 100% load condition and the 110% load condition. Three runs were performed at each location per load condition. Results of the testing are presented in Tables 5-20, 5-21, and 5-22. SO_2 removal efficiency averaged 93.8% for the 75% load condition, 95.8% for the 100% load condition, and 94.6% for the 110% load condition.

With respect to SO_3 emissions at the outlet of the condenser, and therefore H_2SO_4 condensation efficiency, the concentrations for this test series were significantly lower than the previous data. At 75% load, the first two runs were suspected to be contaminated, with SO_3 values of about 82 ppm for each. The third run, a value of 0.44 ppm is consistent with the averages for the 100%

test (0.41 ppm); and the 110% test (1.0 ppm).

Run Number[a]	SO_2 Baghouse Inlet	SO_3 Baghouse Inlet	SO_2 System Outlet	SO_3 System Outlet	SO_2 System Removal Efficiency (%)
	lb/hr (ppmv)				
1	1,203 (2,240)	4.1 (6.0)	59.5 (82)	74.0 (81.6)[b]	95.1
2	847 (1,610)	3.6 (5.6)	52.4 (75)	71.9 (81.9)[b]	93.8
3	1,027 (1,970)	4.1 (6.2)	77.6 (108)	0.4 (0.44)	92.4
Average	**1,026 (1,941)**	**3.9 (5.9)**	- 19 **63.2 (88)**	**48.7 (0.44)**	**93.8**

[a] December 17, 1992

[b] Contamination suspected, not used in average

Run Number[a]	SO_2 Baghouse Inlet	SO_3 Baghouse Inlet	SO_2 System Outlet	SO_3 System Outlet	SO_2 System Removal Efficiency (%)
	lb/hr (ppmv)				
1	1220 (1720)	5.4 (6.2)	49.4 (58)	0.33 (0.31)	95.9
2	1330 (1880)	5.9 (6.7)	58.2 (68)	0.53 (0.49)	95.6
3	1520 (2153)	5.9 (6.6)	61.9 (72)	0.45 (0.42)	95.9
Avg	**1356 (1917)**	**5.7 (6.5)**	**56.5 (66)**	**0.44 (0.41)**	**95.8**

[a] December 18, 1992

Item	Spec.	SNOX
Concentration (%)	93.2	94.7
H_2SO_3 (ppmw)	40	0.4
Iron	50	3.4
Copper	50	0.025
Zinc	40	0.12
Arsenic	1	0.012
Antimony	1	0.002
Selenium	20	0.001
Nickel	1	0.04
Manganese	0.2	0.065
Nitrate	5	2.5
Ammonium	10	2.9
Chloride	10	5.5

5.6 Sulfuric Acid Quality

Sulfuric acid quality is quantified by three main properties - concentration, color, and trace

Run Number[a]	SO_2 Baghouse Inlet	SO_3 Baghouse Inlet	SO_3 System Outlet	SO_3 System Outlet	SO_2 System Removal Efficiency (%)
	lb/hr (ppmv)				
1	1566 (2167)	4.6 (5.1)	95.5 (103)	1.6 (1.4)	93.9
2	1526 (2112)	8.6 (9.6)	73.0 (79)	0.9 (0.8)	95.2
3	134[b] (186)	0.5[b] (0.5)	103.3 (111)	1.0 (0.9)	23.1
Average	**1546 (2139)**	**6.6 (7.3)**	**90.6 (98)**	**1.2 (1.0)**	**94.6**

[a] February 8, 1993
[b] Data not used in average, suspect sample

impurities. Concentration requirements are usually specified as one of several industry standard values, with the most common being 93.2 weight % or 66EBaumeN. This concentration is most common as it represents a minimum freezing point, allowing more convenient transportation modes without freeze protection. Color and impurity limits are more strictly tied to the specific end use of the acid. For instance, an end product of a pigment or synthetic plastic may have color requirements which cannot be satisfied with an off-color sulfuric raw material. In contrast, an ore leaching operation may have no requirement as to sulfuric acid color. In general, however, water clarity is standard.

Limitations on trace impurities in sulfuric acid are dictated in much the same way, by the specifics of the end use. Notably, iron and other metal concentrations are restricted when the intended use of the acid is as an electrolyte for storage batteries. An industry wide standard for sulfuric acid concentration, color, impurities, and the test methods to analyze for each is the U.S. Government Specification O-S-801E (current revision). Individual acid brokers and end users, however, may operate under more or less stringent specifications tailored to their end user or process. As shown by the typical analysis in Table 5-23, the acid from the Niles SNOX plant meets or exceeds the government specifications.

Additional data on acid concentration and acid quantity produced from the SNOX plant are given in Section 4-5.

6.0 ENVIRONMENTAL PERFORMANCE

6.1 Impact on the Environment

With respect to pollution prevention, the ability of the SNOX process to achieve extremely low emission levels was indicated by early performance of the European prototype plants and further verified by the CCII Demonstration Plant performance. Test results obtained during the operations and testing phase of the Demonstration Project indicated that all design parameters were met or exceeded. The plant was designed to achieve the following removal efficiencies: SO_2 - 95%, No_x - 90%, and acid generation with a concentration of 93.2%. Results have typically indicated SO_2 at 95%, NO_x removal at 94%, and acid concentration at 94%; well in excess of U.S. Federal requirements.

Unique design features of the process contribute to these high efficiency levels and contribute also to almost negligible particulate emissions. Because the SO_2 oxidation catalyst follows the NO_x or Selective Catalytic Reduction (SCR) system, anyunreacted ammonia is oxidized. For this reason, higher than typical ammonia feed rates can be employed without concern for ammonia discharge and at the same time producing high removal efficiency of NO_x. With respect to SO_2 removal, collection of sulfuric acid mist has been a problem for previous acid producing technologies. In the WSA tower, the unique employment of small diameter glass tubes, precise temperature control of the condensing process, and individual mist eliminators per tube, SO_2 removal is high and sulfuric acid mist is controlled to less than 5 ppm.

Particulate emissions are impacted by two major features. To minimize dust collection on the semi-molten SO_2 catalyst, a high efficiency fabric filter is the device of choice for a SNOX plant. Teflon coated fiber glass bags are employed (and are being used at the Ohio project) and are capable of emission levels of 1 ng/Nm^3 (.001 lb/MBtu). The remaining dust is collected by the oxidation catalyst and results in almost immeasurable particulate levels. Even without employing a high efficiency fabric filter, the oxidation catalyst removes most remaining dust but must be cleaned more often. Therefore, respirable dust (PM 10) emissions are extremely low from this process.

Another unique feature of the process is its ability to recover additional heat from the flue gas and increase the efficiency of the boiler, by employing the hot air discharge from the WSA condenser as combustion air to the boiler. This additional thermal energy can be used to reduce the coal feed rate or to produce more electricity if boiler heat transfer surface is adequate. In addition to the economic impact of the heat recovery, an increase in the thermal efficiency of a plant results in decreased CO_2 emissions per MW generated. In limestone wet scrubbing systems, reaction of the limestone with SO_2 also liberates additional CO_2. These two factors combine to produce a large net decrease in CO_2 produced by the SNOX process compared to conventional limestone technologies.

The above reductions in emissions were achievable without the production of solid or slurry wastes. Thus the real estate required for a commercial operation is minimal compared to existing wet or dry scrubbing technologies. No large settling ponds or solid waste pile storage are required; no large limestone or other reagent storage is required; and as a result no excessive landfill requirements exist. Therefore the environmental impact on land quality is considered to be neutral.

The impact on water quality is likewise considered to be neutral. Only limited amounts of water are utilized in the system for acid cooling and dilution of the acid produced during operations. Water used for cooling purposes does not directly interact with any chemical process and therefore can be returned to the environment without treatment. Water used for the purpose of dilution is consumed (becomes part of the product acid sold) and therefore does not impact the quality of the water source.

During operations, the Demonstration Project produced marketable commercial quality sulfuric acid. This acid, usable throughout industry, is an asset rather than a waste liability. The revenue from sale of the acid more than outweighs the minimal costs of any on-site storage that may be required, therefore effectively reducing the operating costs of the plant itself.

The quarterly and annual Environmental Monitoring Reports published during the operations and testing phase of the Demonstration Project document the efficiency of the SNOX process. Consistent removals of SO_2 and NO_x at or exceeding design goals were met under various plant operating loads and scenarios, except where planned tests caused reduced efficiency or system upsets occurred. Also, marketable commercial grade sulfuric acid, with a purity within or exceeding design parameters, was consistently produced and sold to an acid distributor.

In summary, the demonstrated benefits of commercial application of the technology are substantial. These benefits are:

- Greatly reduced air emissions without the generation of additional waste streams.
- No alkali reagent required for SO_2 removal
- Salable by-product - sulfuric acid
- Very low particulate emissions, due to additional capture in SO_2 catalyst and WSA condenser
- No increase in CO_2 emissions due to lack of limestone reagent
- Reduction of CO and hydrocarbons
- Furnace integration of recovered heat, and reduction of CO_2 due to increase in furnace efficiency

6.2 Waste Streams and their Disposal

As has been stated, the SNOX process does not produce any substantial waste streams. Three small outputs which can be considered waste streams are additional fly ash, catalyst siftings from cleaning, and spent catalyst.

The amount of fly ash, already a product of the generating plant, will be slightly increased if a high efficiency ESP or baghouse is installed as part of the plant. This additional fly ash capture actually represents a net reduction of particulates released to the air, and therefore constitutes a positive impact to the atmospheric environment. The net increase of fly ash requiring disposal is minimal compared to the ash produced in total by the generating plant.

A mixture of SO_2 catalyst and fly ash will be produced in small quantities whenever the catalyst requires sifting (cleaning). Due to the sticky surface of the catalyst at operating temperature, the catalyst collects a portion of the fly ash which enters the beds. As pressure drop across the beds increases, the catalyst must be cleaned. The siftings are a mixture of catalyst material (vanadium based) and fly ash. At the Niles Demonstration Plant, and in future commercial installations, this material will be sent to a metals reclaimer.

Spent catalyst, from both the SO_2 converter and SCR reactor, will be produced periodically over the life of the system. Catalyst life, estimated by the technology licensor, is expected to range from 8 to 10 years for both. As with the catalyst siftings, both catalysts will be sent to the manufacturer for recycling, or to a reclamation facility for metal recovery.

6.3 Potential Environmental Concerns

Safe storage and transport of ammonia and sulfuric acid are of general concern to prevent unwanted spills and discharges. However, handling of these two materials for the SNOX process is no different than would be followed by other users or producers of ammonia or sulfuric acid. With respect to ammonia, anhydrous ammonia was used at the Niles facility but the process can as easily employ aqueous ammonia, which has a much lower hazard potential. No reportable spills or discharges of either ammonia or sulfuric acid occurred at the demonstration project.

Within the process, ammonia is introduced into the gas stream at the SCR reactor. Any ammonia remaining in the gas stream after reaction with the SCR catalyst is transported into the SO_2 converter, where the ammonia is completely oxidized. Ammonia slip from the SNOX process is thus significantly less than other SCR technologies.

Air toxics testing and analysis was not conducted as part of the SNOX Demonstration Project. Air toxics testing was, however, accomplished as a separate DOE sponsored project (number DE-AC22-93PC93251) by Battelle, Inc. These tests were accomplished in July 1993 and reported in July 1994 under the title, "A Study of Toxic Emissions from a Coal-Fired Power Plant Utilizing the WSA-SNOX Innovative Clean Coal Technology Demonstration", by Battelle

Columbus Operations, Columbus, Ohio. In summary, the report states, "*Removal efficiencies and material balances were calculated for the major and trace elements. Removal efficiencies for these elements were calculated for the SNOX baghouse and for the entire SNOX process. Removal efficiencies for most elements exceeded 99 percent in both cases*."

7.0 ECONOMICS

The economic assessment presented in this section for the SNOX technology is based on a 525 MWe gross power plant, having a net output of 500 MWe before the addition of the SNOX technology. Parasitic power requirements of the SNOX process will alter the power plant's net output. However, because the SNOX process generates and recovers additional thermal energy to be input to the boiler heat cycle, net parasitic load is reduced. This is discussed in detail in the following subsections.

7.1 Economic Parameters

SNOX system costs were developed based on the design premises given in Table 7-1. A typical SNOX plant includes facilities for the following:

- selective catalytic reduction of NO_x
- catalytic oxidation of SO_2 to SO_3
- production of H_2SO_4 in the WSA condenser
- sulfuric acid storage and loading
- ammonia storage
- SO_2 catalyst screening
- balance of plant and auxiliaries

Particulate collection has not been included in the cost of the SNOX plant due to the variability in a retrofit situation. The SNOX process may be able to accommodate the performance of the existing particulate collector, or rely on upgrading of existing equipment to produce a satisfactory particulate inlet for the SNOX plant.

For the case presented, it is assumed that flue gases from the boiler are taken downstream of the air preheater, treated in the SNOX plant, and directed to the stack. The cooling air for the WSA condenser is taken from the atmosphere by means of cooling air blowers and discharged to the boiler air preheater for combustion air. Ammonia, used for the reduction of NO_x, is received at an ammonia off-loading station and stored in ammonia tanks. Sulfuric acid produced by the process is held in storage tanks until transferred to transport vessels through the acid loading station.

Table 7-1 Design Premises for Economic Analysis

ITEM		UNITS	VALUE
Cost of debt		%	8.5
Dividend rate for preferred stock (pre-tax)		%	7
Dividend rate for common stock (pre-tax)		%	7.5
Debt/total capital		%	50
Common stock/total capital		%	15
Income tax rate		%	35
Investment tax rate		%	38
Investment tax credit		%	0
Property taxes and insurance		%	3
Inflation rate		%	4
Discount rate (with inflation)		%	7.925
Discount rate (without inflation)		%	3.744
Escalation of raw materials above inflation		%	0
Construction period		Years	1
Allowance for funds during construction		%	0
Construction downtime		Days	90
Remaining life of power plant		Years	15
Year for cost presented in this report		-	1995
Royalty allowance (based on total process capital)		%	0
Capital charge factor:	Current dollars	-	0.16
	Constant dollars	-	0.124
O&M cost levelization factor:	Current dollars	-	1.314
	Constant dollars	-	1
Power plant capacity factor		%	90.4
Sales tax rate		%	5
Cost of freight for process equipment		%	2
General facilities/total process capital		%	10
Engineering and home office fees/total process capital		%	10
Coal heating value		Btu/lb	11241
Coal ash		%	8.7
Coal sulfur content		%	3.24
Boiler NO_x emission		ppmv	600
SO_2/NO_x removal		%	95/90

7.2 Estimated Process Capital Costs

At the time of assembly of this report, three major sources of capital cost information were available to provide data for a U.S. commercial scale plant: the 35MWe demonstration plant at Niles, OH; the 305 MWe plant in Denmark; and a budget price estimate formulated for a 325 MWe plant in the U.S. While the Danish plant is potentially the best source of data, the fact that it is a European plant limits its usefulness. The demonstration plant has very detailed cost information but a large scale-up is necessary to project to the 300 - 500MWe range. For the budget price estimate, ABBES obtained some firm equipment prices but also estimated much of the equipment from in-house estimating programs developed for other scrubbing systems.

From these three sources of data, ABBES developed a budget estimating procedure based on scale-up factors applied to all balance of plant equipment, including installation, with the process developers (Snamprogetti and Haldor Topsoe) providing a firm cost for their scope of supply. This is the approach that has been employed to estimate the cost of the 500MWe plant (retrofit factor = 1) chosen in this economic evaluation. Therefore, detailed cost sheets of individual components/equipment will not be supplied for this case study.

Area No.	Total Installed Equipment Cost	$\$10^6$	$/kW
100	Raw material receiving and handling system	0.3	0.6
800	SO_2 removal system	22.9	45.8
900	NO_x removal system	15.8	31.6
1100	Flue gas/cooling air handling system	20.7	41.4
1300	By-product processing and handling system	26.5	53.0
1500	Common support systems	3.9	7.8
(A)	**Total Process Capital (sum of process areas)**	90.1	180.2
(B)	General facilities (10% of A)	9.0	18.0
8	Engineering and home office fees (10% of A)	9.0	18.0
(D)	Project contingency (25% of A+B+C)	27.0	54.0
(E)	**Total Plant Cost (A+B+C+D)**	135.1	270.2
(F)	Allowance for funds during construction (3.8% of E)	5.1	10.2
(G)	**Total Plant Investment (E+F)**	140.2	280.4
(H)	Royalty Allowance (0% of A)	0	0
(I)	Preproduction costs (2 months of start up)	2.0	4.0
(J)	Inventory capital	1.5	3
(K)	Initial catalyst and chemicals	8.9	17.8
(L)	**Subtotal Capital (G+H+I+J+K)**	152.6	305.2
(M)	Cost of construction downtime	0	0
(N)	**Total Capital Requirement (L+M)**	152.6	305.2

Using the design premises specified in Table 7-1, the Total Capital Investment for a SNOX plant was estimated and is detailed in Table 7-2. Installed cost of the process equipment is broken down into the $DeSO_x$ area, $DeNO_x$ area, flue gas/air handling, and acid handling and storage. The $DeSO_x$ area includes the SO_2 converters which contain the catalyst for conversion of SO_2 to SO_3, and the WSA condenser modules which condense the H_2SO_4 from the flue gas. In the $DeNO_x$ area, the estimate includes the NO_x reduction catalyst and vessels plus the ammonia storage and injection equipment. Flue gas/air handling equipment is comprised of all flue gas and air ductwork, fans, and the gas/gas heat exchanger used in the regenerative SNOX circuit. Acid handling and storage facilities include the sulfuric acidrecirculation and cooling loop at the outlet of the condensers, water dilution equipment to adjust acid concentration if necessary, and the main storage tanks and loading facilities for transfer of acid to truck or rail. Total Process Capital (TPrC) for this plant is estimated to be $90.1 million.

General Facilities costs are added to the Total Process Capital at 10% ofTprC, Engineering and Home Office Fees at 10% of TprC, and a Project Contingency (reflecting the "preliminary" status of this capital cost estimate) of 25% of TprC plus General Facilities and Engineering and Home Office Fees. Total Plant Cost as shown in Table 7-2 is $135.1 million.

Two years of construction time is assumed for the plant, which results in a construction funds allowance of 3.8%, or $5.1 million. This is added to the Total Plant Cost to produce a Total Plant Investment of $140.2 million.

Royalty Allowance is not included for the SNOX process. License cost is included in the catalyst and component items purchased from thelicensor. Preproduction costs, consisting of two months of start up time, are comprised of fixed and variable operating and maintenance cost, excluding credits generated by the process. These values are taken from Table 7-4, and total $2 million for the two months. Inventory capital is 60 days of variable operating costs, again excluding any process credits, and amounts to $1.5 million. Initial catalyst and chemicals is comprised primarily of the SO_2 and NO_x catalysts, but also includes the initial ammonia charge. This item is $8.9 million. Subtotal Capital, as shown, is therefore $152.6 million. No construction downtime is assumed other than normal outages of the unit, therefore no cost incurred was included for this item. Total Capital Requirement is therefore $152.6 million.

7.3 Projected Operating and Maintenance Costs

Heat addition, transfer, and recovery are of significant importance in the SNOX process as they influence the operating cost of the technology. The process generates recoverable heat in several ways. All of the reactions which take place with respect to NO_x and SO_x removal are exothermic and increase the temperature of the flue gas - NO_x/NH_3 reactions, SO_2 oxidation, SO_3 + water to form gaseous sulfuric acid, and condensation of the sulfuric acid. This heat plus support heat added before the SCR is recovered in the WSA condenser cooling air discharge for use in the furnace as combustion air. Because the WSA condenser lowers the temperature of the flue gas to about 210EF (99EC) compared to the 300EF range for a typical utility plant, additional thermal energy is recovered along with that from the heats of reaction.

	)NO_x or SO_2 (lbmols/hr)	)H (Btu/lbmol)	Heat (MBtu/hr)	MWe
NO_x Reduction	102	176,400	18.0	1.9
SO_2 Oxidation	414	42,200	17.5	1.8
SO_3 Hydration	418	43,000	18.0	1.9
H_2SO_4 Condensation	418	29,700	12.4	1.3
Additional Flue Gas Cooling			129.9	13.7
Sub Total			195.8	20.6
Support Heat			84.6	8.9
Total			280.4	29.5
Plant Heat Input (PHI)			4739	
% of PHI with Support Heat			5.9	
% of PHI w/o Support Heat			4.1	
SNOX Auxiliary Power				12.5

Using the plant design information in Table 7-1, the quantities of recoverable heat were calculated and are shown in Table 7-3. As can be seen, the total heat available is substantial and is equivalent to about 4.1% of the boiler=s total heat input without including support heat which is added before the SCR and SO_2 converter. With this heat included, the total heat available from the SNOX process to the boiler is about 5.9% of plant input. Looking at these values from another perspective, the amount of heat generated by SNOX is equivalent to 20.6 MWe of power without support heat, and 29.5 MWe of power with fuel-fired support heat (using an assumed plant net heat rate of 9478 Btu/kWh). By contrast, the amount of auxiliary power required by the SNOX process is about 12.5 MWe for this size plant and conditions. For a retrofit application, the thermal energy recovered by SNOX can be used to reduce the coal feed rate or to produce more electricity if modifications to the boiler and generating equipment can easily and economically be made.

Of course heat losses in both the flue gas and hot air ducts will decrease the amount of heat that can actually be transferred to the boiler. In addition, not all of the air produced by the WSA condenser can be used under all operating conditions of the boiler and SNOX system. Given these reductions, it is estimated that 70 to 80% of the heat generated by SNOX can be transferred back to the boiler. This transferred heat is available in two forms. About 40% enters the furnace as higher temperature combustion air. The other 60% enters the cycle via the SNOX trim cooler where boiler feed water is used as the cooling medium.

Fixed and variable operating costs were estimated for the 500MWe plant described in Table 7-1, and are shown in Table 7-4. Operating labor is expected to require 1 skilled person three shifts per day and one unskilled person two shifts per day, for a total of 5 skilled man-years and 3.2 unskilled man-years for full coverage. It is assumed that unskilled labor can be coordinated with other unskilled labor requirements for the plant to allow the fractional man-years. Fractional man-years were rounded up in the case of skilled labor. Administrative and support labor was added at 20% of the total labor cost.

Overall maintenance costs, both labor and material, were estimated in accordance with DOE guidelines. A value of 2% of Total Plant Investment was assumed, as for a liquids and gases plant, since the SNOX process is primarily catalytic and the acid collector is also a condenser rather than an absorber (no pumping of acid through the vessel), also reducing maintenance cost.

Power consumption was projected for all auxiliary equipment and totals approximately 12.5MWe for the 500 MWe plant. This was converted to MWh using the capacity factor of 90.7% based on 330 operating days/yr specified in Table 7-1. The remaining consumables and credits were also

Item	$/Unit	Units/Yr	$\$10^6$/Yr
Fixed O&M Costs			
Operating Labor - Skilled	$50k/manyr	5	0.25
Operating Labor - Unskilled	$35k/manyr	3.2	0.11
Administrative/Support Labor	20% of labor		0.07
Maintenance Material	(TPI x 2% x 60%)		1.63
Maintenance Labor	(TPI x 2% x 40%)		1.08
Subtotal Fixed Costs			3.14
Variable O&M Costs			
Power	50/MWh	99,200	4.96
Ammonia	190/ton	7,368	1.40
Natural Gas	2.50/MBtu	672,000	1.68
Cooling Water	0.15/k gal	200,000	0.03
SCR Catalyst	15,000/m^3	23	0.34
SO_2 Catalyst	3500/m^3	114	0.40
Catalyst Disposal	350/ton	86	0.03
Sulfuric Acid	25/ton	173,941	-4.35
Heat Recovery	2.00/MBtu	1,665,576	-3.33
Subtotal Variable Cost			1.16
TOTAL O&M COST (FIXED + VARIABLE)			**4.30**

determined using the 90.4% capacity factor. Ammonia consumption was calculated based on a stoichiometry of 1.02 and 90% NO_x removal efficiency. Support heat before the SCR and SO_2 converter was assumed to be provided by natural gas in Table 7-4, although steam can also

be employed for this purpose. Cooling water demand is that required to cool the sulfuric acid in the recirculation system.

SCR catalyst replacement is based on a single bed life of 10 years. That is, two beds are initially installed and a third is added after 5 years. Subsequently, one bed is then removed and replaced every 5 years. Therefore, each bed is actually in service for 10 years. Yearly cost for the SO_2 catalyst is based on a life expectancy of 10 years and a cleaning frequency of one time per year, with a loss of 2.5% of catalyst per cleaning.

Periodically, spent catalyst will require disposal through a precious metal reclaimer. After the first ten years and every five years thereafter, one bed of spent SCR catalyst will be produced. Every ten years the SO_2 catalyst will be discharged as will the annual siftings of this catalyst. An annualized disposal cost for these wastes is presented in Table 7-4.

Operating cost credits are for the sale of sulfuric acid and heat recovery in the SNOX system for use in the boiler. Determination of sulfuric acid revenue is straightforward with 174,000 tons/year produced and sold at a price of $25/ton. Credit for heat recovery was determined in the following fashion: From Table 7-3, the amount of heat recovered by the SNOX process (280.4 MBtu/hr) is multiplied by a transfer efficiency of 75%. This heat is valued at $2/MBtu, an intermediate value between fuel replacement cost (typically $1.50/MBtu for coal), and what it would be worth if converted to electricity ($4-5/MBtu).

From Table 7-4, subtotal fixed operating and maintenance (O&M) costs are $3.14 million per year, and variable O&M costs are $1.16 million per year, including the credits for sulfuric acid sales and energy recovery. Total O&M, fixed plus variable, is $4.3 million per year.

7.4 Summary of Performance and Economics

Table 7-5 summarizes the performance and economics of CCT processes, including power plant attributes and emissions control data. Performance figures include emissions of SO_2, NO_x, and particulates, both before and after controls. The particulates are given as total suspended particulates (TSP) and/or particulate matter having a diameter of 10 microns or less (PM_{10}). The economics are presented in terms of costs levelized over the projected life of the plant. This calculation is based on the appropriate levelization factors, which are applied to the O&M costs, and a capital charge factor, which is applied to the capital cost.

The capital charge factors and O&M levelization factors have been calculated according to guidelines established by EPRI, taking into account the financial parameters specified in Table 7-1. Two different sets of factors are calculated - one on a current dollar basis which includes the effect of inflation; or a constant dollar basis, which ignores inflation.

Table 7-5 Summary of Performance and Cost Data

Power Plant Attributes	Units	Value
Plant capacity, net	MWe	487
Power produced, net	10^9 kWh/yr	3.84
Capacity factor	%	90.4
Plant life	yr	15
Coal feed	10^6 tons/yr	1.72
Sulfur in coal	wt%	3.24

Emissions Control Data	Units	SO_2	NO_x	TSP	PM_{10}
Removal Efficiency	%	95	90	99	90
Emissions without control	lb/10^6 Btu	5.76	0.69	0.37	
Emissions with control	lb/10^6 Btu	0.29	0.07	0.004	
Amount removed	tons/yr	105,962	11,666	6297	

	Current Dollars		Constant Dollars	
Levelized Cost of Power	**Factor**	**Mills/kWh**	**Factor**	**Mills/kWh**
Capital Charge	0.160	6.36	0.124	4.93
Fixed O&M Cost	1.314	1.08	1.000	0.82
Variable Operating Cost	1.314	0.39	1.000	0.30
Total Cost	-	7.83	-	6.05
Levelized Cost -- SO_2 Basis	**Factor**	**$/ton removed**	**Factor**	**$/ton removed**
Capitol Charge	0.160	230.5	0.124	178.6
Fixed O&M Cost	1.314	38.9	1.000	29.6

Variable Operating Cost	1.314	14.4	1.000	10.9
Total Cost	-	283.8	-	219.1
Levelized Cost - SO_2 + NO_x	**Factor**	**$/ton removed**	**Factor**	**$/ton removed**
Capital Charge	0.160	207.6	0.124	160.9
Fixed O&M Cost	1.314	35.1	1.000	26.7
Variable Operating Cost	1.314	13.0	1.000	9.9
Total Cost	-	255.7	-	197.5

Referring to Table 7-5, the levelized incremental cost of power for a plant using the SNOX process is 6.05 mills/kWh in constant dollars, and 7.83 mills/kWh in current dollars. The levelized cost on a tons of SO_2 removed basis is \$219.1/ton in constant dollars and \$283.8/ton in current dollars. For both SO_2 and NO_x removed, the cost in constant dollars is \$197.5/ton and \$255.7/ton in current dollars.

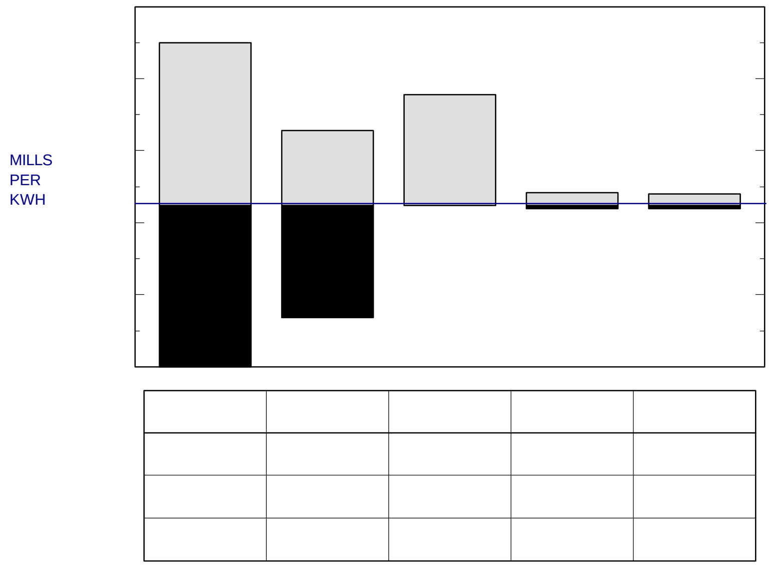
7.5 Effect of Variables on Economics

In order to evaluate the effect of variables on O&M costs, a sensitivity analysis was performed for changes in cost or credit for electric power, SCR and SO_2 catalyst life, heat recovery, and selling price of sulfuric acid. Results of these determinations are shown in Figure 7-1.

In this figure, baseline O&M costs are represented by the horizontal line, at 1.12 mills/kWh. This is the total O&M cost for the case study, the sum of fixed and variable operating expenses on a constant dollar basis. The sensitivity study examines the effect of changes in the stated major variable operating costs on total O&M, independent of each other.

If the sulfuric acid is disposed of with no profit, the operating cost of the system increases to about 2.25 mills/kWh. If it generates a net profit of \$50/ton, O&M costs drop to zero.
For electric power, the cost per Mwh was varied from \$30 to \$70 with the base price at \$50. If electricity cost is raised to \$70, the total O&M increases from 1.12 to 1.64 mills/kWh. Likewise, if power cost is decreased to \$30, O&M costs drop to 0.34 mills/kWh.

Heat recovery refers to the thermal energy recovered by the cooling air in the WSA condenser which is available for use in the boiler. Two cases are listed, one for 75% recovery and the other for no recovery. As stated earlier, the amount of energy generated and recovered by the SNOX



process represents a valuable by-product. Impact on the overall O&M cost by the use of this recovered energy is significant and represents a 0.9 mills/kWh credit for this design case, resulting in total O&M costs of 1.12 with and 2.02 mills/kWh without the heat credit.

With respect to catalyst replacement intervals, both the SCR and SO_2 catalyst lifetimes were varied from their average periods of 10 years each. Neither catalyst life has a significant effect on overall O&M costs even with relatively short life expectancies.

In summary, acid revenue is important but does not make or break the economics of the SNOX process. Heat recovery is also important and should normally be available for incorporation into the boiler heat cycle.

8.0 COMMERCIALIZATION POTENTIAL AND PLANS

8.1 Market Analysis

8.1.1 Applicability of the Technology

As a commercially proven technology, the SNOX process provides an economical and technically superior system for the simultaneous control of NO_x, SO_x and particulate. The competitive capital and operating cost as well as the production of a marketable commodity (H_2SO_4) make the SNOX technology attractive for new and retrofit applications.

A SNOX process installation consists largely of proven, commercially available equipment such as bag filters, blowers, reactors, pumps, etc. The only novel equipment item is the WSA tower, which as previously mentioned, has been tested extensively in European applications, and now in the DOE Demonstration Project.

In summary, commercialization of the technology will be aided by:

- Simultaneous removal of 90+% NO_x, 95% SO_x and essentially all particulate matter.
- Lower per ton SO_2 removal costs for high sulfur coals.
- Lower predicted overall station heat consumption rate from integration of the SNOX unit with the combustion air preheat system. This is due to recovery of the heat released in the SO_2/NO_x reactions and the greater thermodynamic efficiency resulting from lower allowable flue gas stack temperatures once SO_x species have been removed by the SNOX process.
- No increase in solid plant wastes because the SNOX is a non-sorbent process.
- Production of marketable sulfuric acid.

The success of this demonstration has established that the SNOX process is an effective, reliable, and economic approach to the control of the two major pollutants associated with acid rain. Accordingly, this technology has the potential to significantly penetrate the large pre-NSPS boiler market for all design types of boilers (cyclone, stoker and pulverized coal).

Available methods to control NO_x and SO_x are not as effective for use in cyclone fired boilers. Cyclone fired boilers produce relatively low fly ash loadings and therefore, are not very suitable for sorbent injection. Wet scrubbing of SO_2 is viable, but these systems require high capital costs, require large site space requirements, reduce power plant availability, reduce power plant electrical output and increase spent material production. Further, low NO_x burners and overfire or concentric air additions are not compatible with the operating characteristics of the cyclone chamber. Consequently, there is a need for a new technology that is efficient, economical and reliable, and that can be used in retrofit applications.

The SNOX process combines NO_x, SO_x and enhanced particulate matter removal. The system will reduce these emissions while lowering fuel usage, improving station heat rate and producing a marketable by-product, sulfuric acid. Since this process sees only the flue gas, the SNOX technology is applicable to all electric power plants and industrial/institutional boilers no matter what fuel is fired as long as NO_x and SO_x are to be removed. The only limitation is that a moderate amount of space is needed near the boiler flue duct so that the flue gas can be economically transported to the SNOX unit, processed and returned to the stack.

8.1.2 Market Size

While the technology is equally applicable to the new electric power plant market, by far the most important market for the foreseeable future will be the retrofit of existing coal fired power stations.

It is estimated that there are over 1000 electric power generating units representing over 240 gigawatts (GW) of "uncontrolled" coal-fired electric utility capacity in the United States. Of these, there are approximately 410 units representing some 128 GW that are over 100 MW in size or larger, incorporate reheat design, and were placed in service from 1955 to 1975. These units represent the population for which significant investment in plant modifications can be justified so they can continue to produce cost competitive electric power. This unit population also accounts for approximately 60% of the total SO_2 emissions by the electric utility industry. In terms of unit design, they include both wall-fired and cyclone units.

An eight state region, which consists of Pennsylvania, West Virginia, Kentucky, Ohio, Tennessee, Missouri, Illinois, and Indiana, forms an area where many of these units are located. This is also a region where sulfuric acid is produced from elemental sulfur. This region will be the initial focus for commercialization of the SNOX technology.

8.1.3 Market Barriers

As discussed previously, the SNOX technology is applicable to all electric power plants and industrial/institutional boilers regardless which fuel is fired as long as NO_x and SO_x are to be removed - the process "sees" only the flue gas. The only limitation is that a moderate amount of space is needed somewhat near the boiler so that the flue gas can be economically brought to the SNOX unit, processed and returned to the stack.

The following list presents common site-specific factors with respect to power plants and the potential impact on the SNOX process.

Impact of Site Specific Factors on the SNOX Technology

Coal type and characteristics	Two effects exist: 1) Large amounts of arsenic or other catalyst poisons in the coal may reduce catalyst life; 2) Process economics are better for high sulfur coal
Boiler size	No effect other than some diseconomies of scale for small units
Boiler age	No effect
Boiler heat release	No effect
Capacity factor limitation	No effect
Load profile applicability	Only economic effects; economics are better for a base load application

Site Specific Factors (Cont=d)

Boiler firing type	No effect
Boiler firing configuration	No effect
Boiler bottom type	No effect
Geographic applicability	No effect other than feasibility of shipping/ selling the sulfuric acid product
Furnace dimensions	No effect
Reheat or non-reheat steam	No effect
Steam turbine characteristics	No effect
Particulate collector requirements	High efficiency particulate collection required. Higher loadings affect SO_2 catalyst cleaning frequency and replacement interval, impacting operating cost.
Raw materials requirements	Only ammonia for the NO_x removal portion of the process
By-product marketing	Somewhat site specific although sulfuric acid output from SNOX plants will be very small compared to total U.S. production and consumption; will be able to under price "on purpose" production of acid

Items which may be considered potential barriers to the technology include mandatory inclusion of the de-NO_x step in the process, high particulate removal efficiency ahead of the process, and site-specific marketability of the sulfuric acid by-product.

The de-NO_x step is required in the process to prevent oxidation of NO to NO_2, resulting in a brown plume if concentrations are sufficient. Depending on the outcome of federal NO_x regulations, and/or regional controls, some plants may not require NO_x reduction beyond that attainable with combustion controls or in-furnace reactant injection methods. In these cases, the SNOX process will have to compete with technologies which offer SO_2 control alone. It is unlikely that overall operating cost (capital plus O&M) of the SNOX technology will be competitive in these situations.

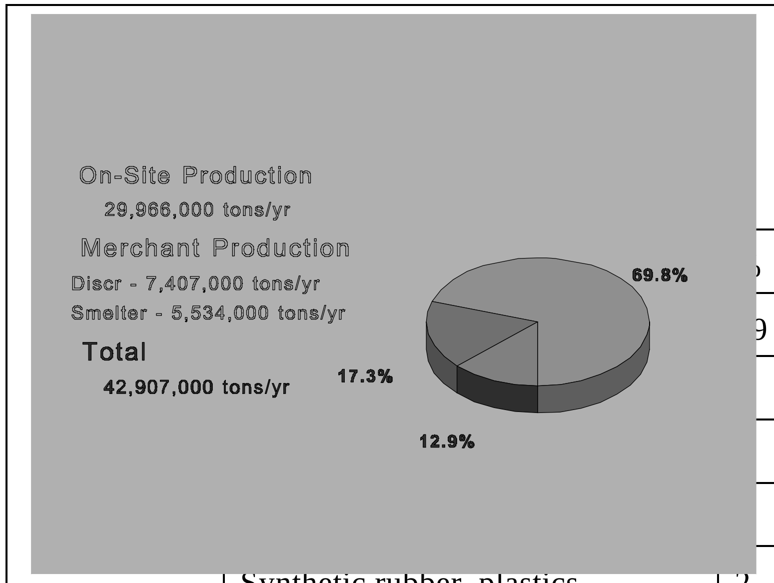
As discussed previously, high efficiency particulate removal is not a necessity but does lower operating costs due to its impact on SO_2 catalyst cleaning frequency (labor), make-up catalyst costs for losses during cleaning, and lower catalyst life. Some retrofit situations may lack space for particulate equipment upgrades, and the SNOX operating cost may be too expensive with the existing particulate devices.

With respect to sulfuric acid, the overall U.S. market is large with respect to the acid output from a typical 500 MWe plant using the SNOX process, but regional use varies and the cost of transportation will become significant if users are located a considerable distance from the plant. As discussed previously, the quality of acid from a SNOX plant meets U.S. specifications and will therefore be readily marketable. Size and characteristics of the sulfuric acid market are discussed in the following section.

A local acid supplier, PVS Chemicals, was contracted to purchase and distribute the acid from the Demonstration Plant. PVS is a large regional marketer and producer of sulfuric acid serving the industrial Midwest in New York, Ohio, Michigan and Illinois. The acid is being sold primarily to the agricultural industry and to the steel industry for pickling.

8.1.4 U.S. Sulfuric Acid Market

According to a recent estimate, approximately 44.2 million tons of sulfuric acid were used in the U.S. during 1991. As indicated in Table 8-1, phosphate fertilizers, ore processing, and inorganic chemicals account for 80% of the total consumption. The remainder is used by a wide variety of industries, including refining, paper, rubber and plastics. Total production of sulfuric acid in the U.S. in 1991 was approximately 46.2 million tons. This represents an operating load of about 96%. Annual growth in the U.S. sulfuric acid market has historically been 0.9% but is expected to be only 0.5% over the next few years.



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Synthetic rubber, plastics	2
Pulp and paper	2
Organic chemicals	2
Other	11
Total	100

The market is very price competitive and suppliers are utilizing various forms of price related incentives such as volume discounts in order to encourage sales. Many large volume suppliers are located near their end-users in order to minimize transportation costs. In fact, 70% of the acid currently produced is used captively at the production point (See Figure 8-1, On-Site Production). Because of increased by-product acid production in the 1980's, many plant closings have occurred since that time period. This has helped bring production levels more in line with end-user demand. In addition, recent increases in fertilizer exports and in the use of sulfuric acid in mineral leaching operations, particularly copper, have helped to stabilize the market even further.

The majority of sulfuric acid produced in the U.S. is from the burning of sulfur in the "contact" process, which accounts for the use of approximately 84% of the elemental sulfur supply. Sulfuric acid generated with this process, known as virgin acid, represents some 87% of the total acid produced. As shown in Figure 8-1, on-site production and use of virgin acid was 29.97 million tons per year, and virgin acid manufactured for sale was 7.4 million tons - a total of 37.37 million tons per year of virgin acid production. Since the 1980's, government regulations have required smelters to treat their off gases for recovery of sulfur compounds. This has resulted in an additional supply of commercial sulfuric acid which has steadily grown to represent the remaining 13% of the current market - 5.53 million tons in Figure 8-1. Merchant production is the total of that produced for sale and is the sum of the 7.4 million virgin tons plus 5.53 million tons of smelter acid, or 12.93 million tons per year.

Although it is possible that SNOX acid or other recovered acid could displace on-site production,

Plant Size (MWe)	385	Load Factor (%)	85
Firing Rate (tph)	152	SO_2 Removal (%)	95
Fuel Heating Value (Btu/lb)	11,750	NO_x Removal (%)	
Sulfur Content (%)	2.9	SNOX	95
NO_x Emission (lb/MBtu)	0.7	WFGD+SCR	80
Fuel Ash (%)	0.25	Capital Charge Rate (%)	15

it is unlikely and the more realistic market for SNOX acid is the 12.93 million tons per year of merchant production. A typical 500 MW e plant, burning 3% sulfur fuel and employing the SNOX process, can recover about 107,000 tons per year of concentrated sulfuric acid. This quantity, however, represents only 0.83% of the U.S. merchant production market. Thus, a substantial number of utility sulfur recovery plants would be necessary in order to significantly affect the market.

8.1.5 Economic Comparison with Competing Technologies

The most likely competing technology, considering the SNOX multi-pollutant capture capability, would be the currently typical limestone, forced oxidation wet flue gas desulfurization (WFGD) system, for the control of SO_2, with an SCR for the control of NO_x. A case study was made and presented at the EPRI/EPA/ DOE SO_2 Control Symposium in Miami, Florida, in March of 1995. A summary of the case study is provided herein and effectively demonstrates the economic viability and competitiveness of SNOX system compared to available technologies.

Capital and operating costs were developed based on the design premises shown in Table 8-2. The study case assumes a conversion to a low cost, high sulfur fuel (hydrocarbon emulsion) in a location requiring SO_2 and NO_x control, upgraded particulate collection, and minimization of water plume formation. The WFGD system is designed for salable gypsum production, and both the SNOX and WFGD cases include particulate control via an ESP.

SNOX system costs were developed based on the design premises stated and the following scope:

- selective catalytic reduction unit for treatment of NO_x

- catalytic oxidation reactor for conversion of SO_2 and SO_3
- H_2SO_4 condensing towers (WSA condenser)
- gas/gas heat exchanger
- heat exchanger for boiler feedwater heating
- sulfuric acid storage and loading facilities
- ammonia storage and injection equipment
- SO_2 catalyst screening equipment
- balance of plant and auxiliaries

For both systems, flue gases from the boiler are taken downstream of the air preheater, treated in the flue gas cleaning plant, and directed to the stack. Ammonia, used for the reduction of NO_x, is received at an ammonia off-loading station and stored in ammonia storage vessels. Sulfuric acid or gypsum produced by the processes is stored in tanks or stacks until transferring to transportation vessels.

For the WFGD/SCR system, costs are based on the following scope.

- selective catalytic reduction for treatment of NOx
- boiler air heater modifications
- limestone unloading and processing equipment, including wet ball mill system
- single open spray tower design SO_2 absorber with auxiliaries
- primary and secondary dewatering equipment to achieve salable gypsum by product

! gypsum storage and loading facilities
! gas/gas heat exchanger
! wastewater treatment system
! balance of plant and auxiliaries

Capital costs for both the SNOX system and the WFGD/SCR system were developed and are listed in Table 8-3. These costs are broken down into the subsystems shown. (Total Plant Cost for SNOX is comparable to line E of Table 7-2, with a lower project contingency)

	SNOX	WFGD /SCR
DE-SOx Area	46.0	51.2
DE-NOx Area	14.8	17.5
Air HeaterMods.		3.5
ESP	8.5	6.7
Gas/Air Handling	17.2	9.6
By-Product Area	4.1	6.8
Reheater		7.0
Total	90.6	102.3
Total ($/kW)	235	266

Some differences in cost for the same item exist between the SNOX plant and the WFGD/SCR plant. For the SCR, it is estimated that the WFGD SCR will be more costly due to its location in the 'dirty' gas steam, where larger catalyst channels are necessary to handle the fly ash. Soot blowers are also required. For the SNOX clean side SCR, smaller catalyst modules are required, no soot blowers, and somewhat less catalyst surface area due to operating at higher ammonia stoichiometries.

With respect to the ESP, the SNOX application is more expensive due to its design for an outlet of 5 mg/Nm^3 compared to 10 mg/Nm^3 for the WFGD/SCR system. This high efficiency ESP is employed to minimize the frequency of cleaning for the SNOX SO_2 catalyst, due to the catalyst removing 90 - 95% of the particulate which enters.

Fixed and variable operating costs were estimated for the 385MWe plant described in Table 8-2, and are shown in Table 8-4.

	Cost	SNOX k$/year	WFGD/SCR k$/year
Operating Labor		362	575
Maintenance Labor & Material		382	765
Electricity	$.03/kWh	2105	1862
Natural Gas	$2.50/MBtu	1112	
Limestone	$15/ton		1500
Ammonia	$150/ton	537	452
Water	$.50/k gal	18	105
SCR Catalyst (including disposal)		209	433
SO_2 Catalyst (including disposal)		378	
Sub Total Costs		**5103**	**5692**
Sulfuric Acid	$10/ton	(955)	
Gypsum	$5/ton		(922)
Heat Recovery	$1.50/MBtu	(1935)	
Sub Total Credits		**(2890)**	**(922)**
Net Total O&M		**2213**	**4770**
Net Total O&M (mills/kWh)		**0.77**	**1.66**

Operating labor for the SNOX plant is based on two operators per shift, while the WFGD/SCR plant employs three people per shift. The difference is due to the more labor intensive aspect of operating a wet FGD plant producing gypsum. Overall maintenance costs, both labor and material, were estimated for WFGD/SCR from actual operating information. For SNOX, the estimate was determined from available commercial data along with the assumption that maintenance will be less than WFGD due to much less rotating equipment.

Power consumption and all other consumables were projected for both systems using the plant availability factor of 85%. Water usage for the SNOX process is comprised of a small amount of acid dilution water plus cooling water to cool the acid after it exits the condenser. For the WFGD/SCR system, makeup water is required for evaporative losses in the flue gas and the small amount of water which is contained in the gypsum.

SCR catalyst replacement for the SNOX system is based on a single bed life of 10 years. That is, two beds are initially installed and a third is added after 5 years. Subsequently, one bed is then removed and replaced every 5 years. Therefore, each bed is in service for 10 years. Yearly cost for the SO_2 catalyst is based on a life expectancy of 10 years and a cleaning frequency of once per year, with a loss of 2.5% of catalyst per cleaning. The SCR life for the WFGD/SCR plant is assumed to be one-half of the life of the clean side SNOX SCR, or five years. Periodically, spent catalyst will require disposal through a precious metalreclaimer. These costs are included in the SO_2 and SCR annualized costs.

Operating cost credits are for the sale of sulfuric acid or gypsum (estimated credit per/ton is shown in Table 8-4), and heat recovery in the SNOX system for use in the boiler. Details of the heat recovery are discussed in Section 7-3.

Referring to the Net Totals in Table 8-4, the O&M costs for the 385 MW plants are projected to be 1.66 mills/kWh for WFGD/SCR and 0.77 mills/kWh for SNOX. Total operating cost for the processes is calculated from the capital cost and the yearly O&M costs. Although other financial items may typically be added by the end user to the capital cost developed for the two systems, the approach taken provides a relative comparison between WFGD/SCR and SNOX for the application examined.

	SNOX (k$)	WFGD/ SCR (k$)
Capital Charges	13,590	15,345
First Year O&M	2,213	4,770
Total Operating Cost	15,803	20,115
Mills/kWh	5.51	7.02

Referring to Table 8-5 yearly capital charges are calculated based on the 15% assumed rate from Table 8-2. These values are added to the first year O&M costs and provide a relative first year total operating cost for the two systems.

Another item of consideration with the SNOX process is the effect of inflation on O&M costs over the life cycle of a project. Compared to high O&M processes, thelevelized cost of SNOX will benefit from lower absolute values of O&M costs per year for a given inflation rate.

8.2 Commercialization Plans

A team approach will be utilized to market the SNOX technology. The marketing effort to implement this technology throughout the coal fired utility industry will be conducted by the Team of ABB and Snamprogetti USA. These two team members are among the most substantial companies in the world and the only companies to have experience with this technology.

As stated earlier, an eight state region, consisting of Pennsylvania, West Virginia, Kentucky, Ohio, Tennessee, Missouri, Illinois, and Indiana, forms an area where approximately 128 GW of existing coal fired electric stations are located that have the potential for conversion to the SNOX technology. This region will be the initial focus for commercialization of the SNOX technology.

Since the inception of the project in the 1988-89 time frame, numerous requests for technical and cost information, budget proposals, and fixed-price proposals have been received from electric utility operators. Although no U.S. sales have been made at the present time, utility perception of the viability of the technology has been positive and has not been a deterrent to selling the process. Visitors to the SNOX Demonstration Plant in Niles, Ohio have been impressed with the system's simplicity, cleanliness, reliability, and overall particulate/NO_x/SO_2 removal performance. Many of the interested parties have also visited the Danish 305MWe plant and were likewise impressed with its appearance and operation.

In addition to technical performance, minimal operating cost, resulting mainly from the lack of an SO_2 reagent, salable by-product, and recovery of thermal energy, has also been attractive to potential customers. At a time when utilities are again looking at waste heat recovery, the SNOX process has a built-in heat recovery and integration scheme. Although capital cost of the SNOX process is higher than conventional technologies, total operating cost (including O&M) is lower for most situations.

Part of the reason for lack of sales to date is related to the size of the actual market which occurred in compliance with Phase I of the Clean Air Act Amendments. This market was much smaller than predicted for various reasons and therefore did not offer a large "pool" of potential flue gas cleaning sites. Also, because of the limited degree of NO_x regulation under Phase I, utilities were not forced to select NO_x control processes with reduction capabilities superior to conventional technologies.

A recently developing market niche is the conversion by oil-fired plants to low cost, high sulfur, hydrocarbon emulsion fuels. Even with the addition of pollution control equipment, the fuels are priced to produce competitive power rates. Typically these are eastern U.S. plants which have port access and are in major cities, and because of their location many are in non-attainment areas with respect to ozone (and one of its precursors, NO_x). Therefore NO_x control is required as well as SO_2, which enhances the competitiveness of the SNOX process. There opportunities are being actively pursued at the present time.

In summary, the U.S.-wide NO_x regulations which are forthcoming under Phase II will provide an impetus for the affected utilities to examine combined NO_x /SO_x control technologies such as the SNOX process. Air toxics control (along with PM10) is likely to be important in the near future and the performance of the SNOX process in these areas will be a major selling point. Utilities are also re-examining the scenario of low-sulfur coal, low NO_x burners and upgraded particulate collection due to the amount of boiler tuning that may be necessary for satisfactory operation. It is definitely less troublesome and may be more economic in the long term to employ post combustion NO_x control instead of furnace modifications.

9.0 CONCLUSIONS AND RECOMMENDATIONS

The SNOX Demonstration Project met its design objectives of:

- Demonstrate NO_x and SO_2 removals of 90 and 95%, respectively
- Demonstrate the commercial quality of the product sulfuric acid.
- Satisfy all Environmental Monitoring Plan requirements.
- Perform a technical and economic characterization of the technology.

Sulfur dioxide removal was consistently in the 95 to 96% efficiency range, and nitrogen oxides removal exceeded its target value by 3 - 4 percentage points, typically being 93 to 94% efficiency. With respect to sulfuric acid quality, its concentration and composition have met or exceeded the requirements of the Federal Specification for Class 1 for species analyzed. Commercial grade acid is specified as 93.2 wt. %, and the demonstration plant acid was consistently in excess of this value. A local acid supplier purchases and distributes the acid from the Demonstration Plant.

Environmental Monitoring Plan requirements were met throughout the project, and the technical and economic characterization of the technology are contained in this report.

With respect to commercialization, it can be stated that the need for multi-pollutant capture will drive application of the SNOX process. Nitrogen oxides control at a high efficiency must be required to justify the SCR control provided by the SNOX process, thereby competing with other processes involving SCR. The SNOX process's very low emission rate of NO_x, SO_x, particulate (including PM10), and hydrocarbons make it an ideal candidate for non-attainment areas.

Heat recovery is an important component of the overall low operating cost of the system. While the integration of this recovered heat into the boiler cycle has been accomplished efficiently in the Danish 300 MWe unit, other methods or concepts may need to be employed for boiler-specific applications in the U.S. In addition to development in this area, capital cost reduction is always of importance and will be examined as the process is proposed for various applications.

Utilization of the process in a "hot scheme" wherein particulate removal is accomplished at SCR and SO_2 reactor temperatures is also an area which requires further development, primarily with respect to the high temperature particulate filtration. In this scheme, the flue gas temperature is not lowered until the end of the SNOX process, and the reheat loop and associated heat exchanger are eliminated. A decrease in capital and operating cost would be realized.

APPENDIX A - Data Collection Plan - Revision 5

APPENDIX B - Operating Data